ECONOMIC ANALYSIS OF A NEW GAS TO ETHYLENE TECHNOLOGY

A Thesis

by

ALI ABDULHAMID ABEDI

Submitted to the Office of Graduate Studies of
Texas A&M University
in partial fulfillment of the requirements for the degree of

MASTER OF SCIENCE

May 2007

Major Subject: Chemical Engineering

ECONOMIC ANALYSIS OF A NEW GAS TO ETHYLENE TECHNOLOGY

A Thesis

by

ALI ABDULHAMID ABEDI

Submitted to the Office of Graduate Studies of
Texas A&M University
in partial fulfillment of the requirements for the degree of

MASTER OF SCIENCE

Approved by:

Chair of Committee, Kenneth. Hall
Committee Members, John Baldwin
Maria Barrufet
Head of Department, N.K. Anand

May 2007

Major Subject: Chemical Engineering

ABSTRACT

Economic Analysis of a New Gas to Ethylene Technology. (May 2007)

Ali Abdulhamid Abedi,

B.S., The University Of Texas at Austin

Chair of Advisory Committee: Dr. Kenneth R. Hall

Ethylene is one of the most important petrochemical intermediates and feedstocks for many different products. The motivating force of this work is to compare a new process of ethylene production developed at Texas A&M University to the most common processes. Ethylene is produced commercially using a wide variety of feedstocks ranging from ethane to heavy fuel oils. Of them, the thermal cracking of ethane and propane using a fired tubular heater is the most common process in the United States. In Europe and Japan, where natural gas is not abundant, thermal cracking of naphtha using a fired heater is the most common process. In addition to these processes; ethylene could also be produced from crude oil by autothermic and fluidized bed techniques and from coal and heavy oils by synthesis from carbon monoxide and hydrogen.

At Texas A&M University, a group of researchers developed a new process that can convert natural gas into liquids (GTL) or to ethylene (GTE). This technology is a direct conversion method that does not require producing syngas. When selecting a process for ethylene production, the dominant factor is the selection of hydrocarbon feedstocks. Based upon plant capacity of 321 million pounds of ethylene per year, this study has shown that using natural gas, as a feedstock, is more economical than using ethane, propane, naphtha, and other feedstocks. Therefore, it is more economical to

convert natural gas directly to ethylene than separating ethane or propane from natural gas and then converting it to ethylene. A process simulation package ProMax is used to run the GTE process; and a software program, Capcost, is used to evaluate fixed capital costs of the GTE process. Finally, the cost index is used to update the cost of the other processes of ethylene production today.

DEDICATION

TO

MY PARENTS

ACKNOWLEDGMENTS

All the praises are due to the God, the one, the Creator, All Mighty, the Most Beneficent and Most Merciful. Deep affection is especially due to my parents who supported and encouraged me throughout my study.

I would like to express my gratitude to my advisor, Professor Kenneth R. Hall, for his support and guidance throughout this research. His continuous supervision is one of the most important factors that drive this research work to excellence. His dedication to share his time and expertise is immensely appreciated.

I would like to extend my appreciation to my committee members, Dr. John Baldwin and Dr. Mariah Barrufet, for serving on the defense committee.

I would like to extend my appreciation to those who helped me through out my master study, especially Dr. Mahmoud El-Halwagi, for his support and encouragement. His time and effort are greatly appreciated.

Special thanks are extended to Dr. Abdoljalil Varavei for his support and helpful comments. Thanks to my colleagues Mr. Saquib Ejaz, Mr. Mert Atilhan, and Mr. Diego Cristancho.

TABLE OF CONTENTS

Page

LIST OF FIGURES

Page

Page

LIST OF TABLES

Page

1. INTRODUCTION

1.1 Background

Ethylene, in the Geneva nomenclature “ethene” ($H_2C=CH_2$), is the lightest olefinic hydrocarbon. At ambient condition it is a colorless, flammable gas of the same density as nitrogen with a slightly sweet odor. Ethylene does not freely occur in nature and yet represents the organic chemical consumed in the greatest quantity worldwide.[1] It is a basic raw material for a large variety of industrial products, either directly as in the production of polyethylene, or after further reaction with other chemicals as in the production of polyvinyl chloride, polystyrene, and polyester.

Ethylene is one of the most important and largest volume petrochemicals in the world today. It is used extensively as a chemical building block for the petrochemical industry. Worldwide demand for ethylene has grown steadily in the past and is expected to reach 140 million tons by the year 2010.[2] The important of ethylene results from the double bond in its molecular structure that makes it reactive. Ethylene can be converted industrially into a variety of intermediate and end products.[1] The major use of ethylene is conversion to low-and high-density polyethylenes, which are used in such applications such as construction, communications, packaging, and manufacturing of industrial and domestic products.

This thesis follows the style and format of *AIChE*.

Other significant uses of ethylene include chlorination to ethylene dichloride, used in the manufacture of the polyvinyl chloride (PVC), oxidation to the ethylene oxide, an intermediate in the manufacture of polyester fibers and films, and the conversion to ethylbenzene, an intermediate in the manufacture of polystyrene.[3] According to The Innovation Group (TIG), a management consulting service, polyethylene is the most prominent product that uses ethylene as a feedstock.[4] Figure 1 presents major ethylene products based upon the TIG study.

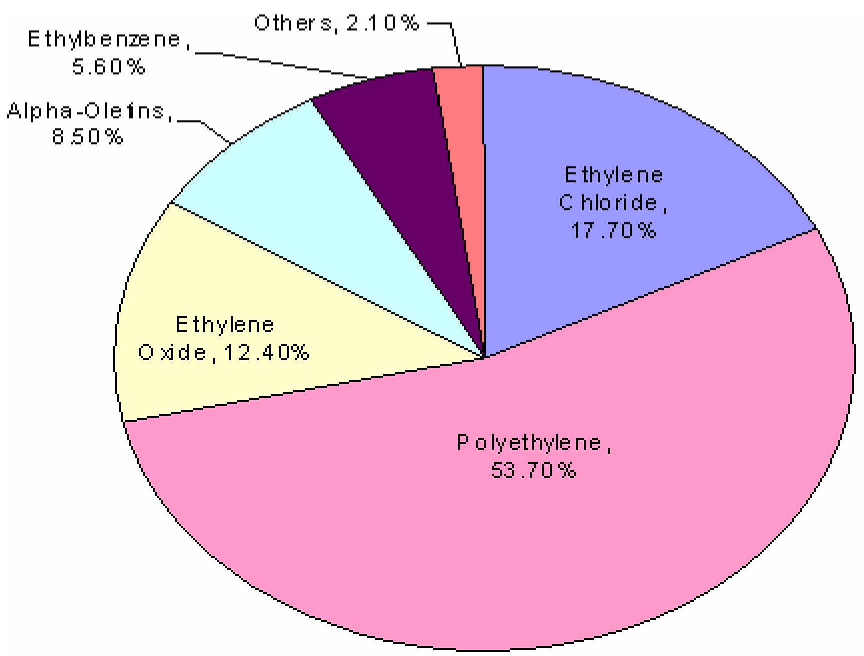

Figure 1: Major ethylene uses in industry.

Since the late 1930's a new industry has become into being in the United States that has gradually taken root and expanded in all of the industrialized countries throughout the world. Before that time ethylene has been recovered in relativity small quantities from coke oven gas as a by product in operations to produce metallurgical coke for the steel industry.[1] Ethylene has become an important industrial intermediate and various technologies have been utilized in ethylene production. Recently, ethylene has taken the place of acetylene in virtually all large-scale chemical synthesis. However, acetylene itself is a byproduct of modern ethylene production.[5]

Ethylene products from hydrocarbons consist basically of four operations: the thermal cracking, quenching, gas compression/treatment, ethylene purification, and refrigeration. In the thermal cracking, several types of pyrolysis reactors have been proposed and commercialized. These pyrolysis reactors include (1) direct heating (2) indirect heating (3) autothermic cracking and (4) others.

Ethylene production is based upon thermal cracking of hydrocarbon feedstock, while ethylene purification is based on compression and refrigeration. The first step in ethylene production is thermal cracking of the hydrocarbon feedstock. Thermal cracking of natural gas liquids (NGL) or crude oil fractions in the presence of steam is still the dominant method for the production of ethylene. According to the Chemical Market Association, Ins., ethane and naphtha are the main feedstocks used to produce ethylene in North America. Figure 2 shows the most common feedstocks used for ethylene production in North America in 2000.

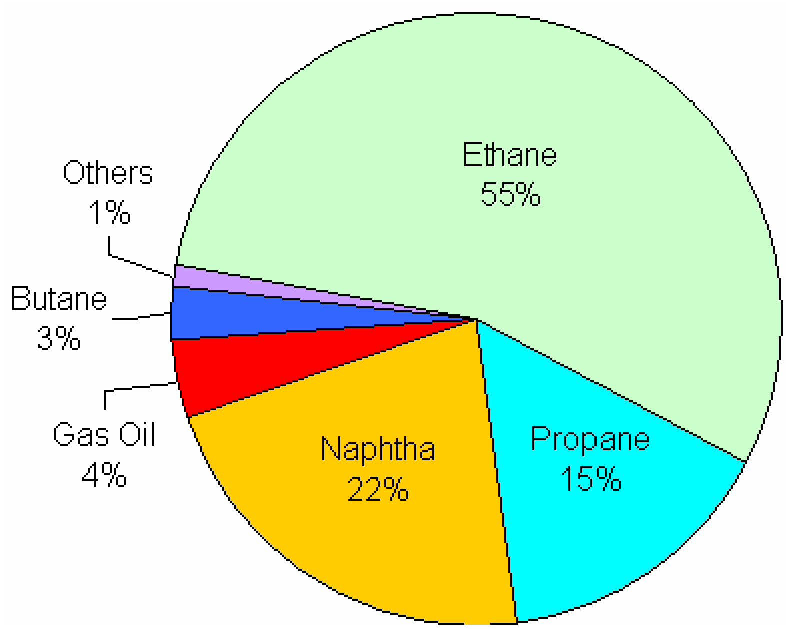

Figure 2: Ethylene production by feedstock in North America 2000.

This thermal decomposition results from adding heat to the feed to break its chemical bonds. The steam does not enter directly into the reaction, but it enhances the product selectivity and reduces coking in the furnace coils. The product of this thermal cracking process is a mixture of hydrocarbons, which extends from hydrogen and methane to gasoline and gas oil.[6]

In general, every kind of hydrocarbon, including paraffins, olefins, naphthenes, and aromatics, can be a starting material for the production of ethylene by thermal cracking. In actual practice, economics considerations narrow the choice of starting materials. The thermal cracking process is essentially one of the dehydrogenations, accompanied to some extent by polymerization and reactions between products to form the ring structure of aromatics and naphthalenes. Generally speaking, the product complexity increases markedly as the molecular weight of the feedstock increases.[7]

Most current ethylene processes are basically similar to each other. Ethylene plants use similar separation units. The purification sequence of ethylene begins with the quenching process. The products that exit the cracker are quenched rapidly in transfer line exchangers that generate high-pressure steam. The heavy oil fraction is separated from the lighter fraction by running the furnace effluent in a quenched tower. At this stage, acid gases are removed from the cracked gas via a caustic wash operation. Then the compressed cracker gas is then dried and chilled in the chilling train and fed to the recovery process.[6]

In the recovery process, a chain of distillation columns, vessels, and heat exchangers utilizing refrigerants as well as cold water and steam are used. In this process, all of the different components are recovered. Usually, in the fired tubular heater processes, ethane and propane are recycled back to the thermal crackers. One problem in the recovery process is components that have similar molecular weights and associated low relative volatilities. These components are very difficult to separate using distillation. Distillation columns to separate these components are the most expensive and intensive units in ethylene plants, because the distillation requires high energy consumption, increased refrigeration capacity and a large number of stages, all of which increase both the capital and production costs.[6] Byproducts are also recovered at this stage. The byproduct yield depends upon the nature of the hydrocarbon feedstock. As the molecular weight of the feedstock increases, ethylene yield decreases and byproducts yield increases. This increase the capital costs of the separation section. Propylene is the most valuable byproduct in ethylene production.

In recent times, researchers have developed some novel technologies to replace the thermal cracking process in ethylene production. These processes use catalysts to convert methane or methanol to olefins, or to perform hydrocarbon pyrolysis using heterogeneous catalytic system. Because of the low yield and limited capacity, these processes do not have significant application in the ethylene production industry.[8, 9] Therefore, steam and thermal cracking technologies still are the best for achieving high yields of ethylene from available feedstocks.

Recovery, separation, and purification of the cracked product are of great importance in ethylene manufacture. They are important not only from a cost standpoint, but also to produce a product with satisfactory purity for its intended use. In addition to the various kinds of hydrocarbons leaving the thermal cracker, the pyrolysis gases contain small amounts of contaminants such as acetylene, acid gases, dines, coke, and hydrocarbon polymers. The increased demand of ethylene for polyethylene and ethylene oxide by direct oxidation processes requires an ethylene purity of 99.9% or higher. The major recovery and purification steps for ethylene are quenching of the conversion products of the furnace followed by compressing the process gas, removing of acid gas such as hydrogen sulfide, removing of water before the separation of ethylene, removing of acetylene and acetylenic materials from the process gas, and separating methane and lighter materials followed by separation of ethylene.[7]

Ethylene production grew significantly in the 1950’s and 1960’s, because of abundantly available and inexpensive light hydrocarbon feedstocks such as ethane and propane derived from natural gas and naphtha form crude oil. Given technology advances in constructing increasingly larger and more economical production facilities in the early

1970's, ethylene costs have steady declined spurring consumption of ethylene in the production of its derivatives. In the past ten years, ethylene demand and price have fluctuated based upon the economical growth in the United Stated and the rest of the industrial world. As seen in table 1, ethylene demand increasing from 1997 to 2000 and deceased from 2000 to 2002.

Table 1: Historical data of ethylene demands and price from 1997 to 2002.

Year	Demand (million pounds of ethylene)	Price (¢ per pound)
1997	55,632	25.90
1998	57,858	20.20
1999	61,110	22.50
2000	62,753	31.65
2001	55,810	29.50
2002	58,915	23.90

More recently, the global ethylene market has grown at an estimated 5% because of strong economical growth throughout the world. Total ethylene consumption for 2006 should reach more than 110 million metric tons. Chemical Marketing Association CMAI experts predict that over the next five years, world ethylene demand will grow over 4%. In addition, global ethylene capacity utilization declined in 2001 through 2003 because of low demand growth. However, operation rates increased in 2004 through 2006 because of resurging ethylene demand at a time of limited capacity. The global utilization rate should remain high in 2007 and 2008, but a sharp decline in global capacity utilization by of the end of decade is expected according to CMAI.[9]

The US position as the low cost ethylene manufacturer has been eroded by large ethylene installations based upon low cost feedstock in the Middle East. At the same time, US exports have declined dramatically because of growing Asian self-sufficiency.

Between 1997 and 2002, US ethylene imports have increased from 4.8 billion pounds to 10 billion pounds, while exports have declined from 240 million pounds to 70 million pounds.[4]

Although many economic uncertainties surrounding the petrochemical industry, ethylene production and consumption should grow because continuing replacement of natural and inorganic materials with organic synthetics and the further development of radically new synthesis materials.[1]

1.2 Motivations

Ethylene is an essential component in the manufacture of many industrial and consumer grade products. Chemical companies have a variety of options for feedstock as well as processes to produce ethylene. Economics and environmental issues are the dominant factors considered in the choice of feedstock and processes of ethylene production.

Throughout the history of ethylene production, many process modifications have been instituted because of increased energy or environmental constraints. Investigations were necessary to determine the effect of these new technologies or modifications.

In the past forty-five years, there have been some improvements and advances within the conventional ethylene production technology. In thermal cracking, researchers worked on increasing product yield, feedstock flexibility, and thermal efficiency. In purification and recovery, there has been progress in different unit operations such as in distillation, refrigeration, and separation.[6]

Most studies in the literature related to ethylene have sought to improve the current process technology, particularly to decrease the energy requirements. Most of the recent studies in ethylene production focus upon reduction of both production and capital costs as well as separation efficiency. For instance, a study by KTI (Kinetics Technology International) concludes that the front demethanizer together with more back-end acetylene hydrogenation could be optimized.[10]

While there have been many studies in the literature to improve the conventional ethylene process, the effort of this thesis has concentrated upon utilizing new technology that has not been evaluated economically for ethylene production. The new technology of ethylene production is gas to ethylene, (GTE), which was recently developed at Texas A&M University. This technology uses a clean and abundant feedstock to produce ethylene. A literature search showed that most of the new approaches developed to convert hydrocarbons to ethylene exhibited deficiencies both technically and economically when compared to the widely used steam thermal cracking technology. Therefore, I intended to do an economical analysis of GTE technology and compare it with the existing technologies to see if GTE can produce cheaper ethylene than other technologies.

In order to study extensively the effect of GTE in an ethylene market, simulation is necessary to analyze the process using accepted economic measures. The environmental and energy constraints are also considered in the final design.

The interest in natural gas as an alternative feedstock for ethylene production stems mainly from its clean burning qualities, its domestic resource base, and its commercial availability. As the price of crude oil increases and an oil shortage looms, in the future, it becomes a concern for scientists trying to use natural gas as an alternative source of energy and as a feedstock in chemical industries. In fact countries that have a large supply of natural gas have started investing on research in this area. New technology is being developed and applied to make better use of natural gas. Liquefied natural gas (LNG) and gas to liquid technology (GTL) are hot areas for investigation and research in some countries like Qatar. The key influences on the competitiveness of these projects are the capital costs, operating costs, natural gas costs, scale and ability to achieve high production rates. For example, GTL not only adds value, but also can produce product that could be sold or blended into refinery stock as a superior products with less pollutants.

In this work, most commercialized process of ethylene production was identified and gas to ethylene technology (GTE) was recognized as a promising alternative to replace old processes. This technology has been selected for an economical study to determine the possibility of replacing or modifying other ethylene processes. For this reason an economical comparison was made between GTE and other ethylene production processes. Although economical evaluation of GTE process has been done on small-scale plant, it has not been compared to other traditional methods of ethylene production to determine the profitability and efficiency of the GTE process.

1.3 Objectives

The main purpose of this study is to estimate the cost of ethylene production using natural gas as a feedstock and to make an economic comparison between gas to ethylene technology and other methods used to produce ethylene. Ethylene is produced commercially by three different methods: fired tubular heater using ethane, propane, and naphtha, autothermic cracking using crude oil, and synthesis from hydrogen and carbon monoxide. Ethane and propane are usually extracted form natural gas and then converted to ethylene. Our goal is to see if producing ethylene directly from natural gas is more economical than extracting ethane and propane from natural gas and then converting to ethylene. We also want to compare GTE, to autothermic cracking and syngas reformation. GTE technology is a new approach to produce ethylene, whereas the other methods have been commercialized for a long time. In any chemical process, when different feedstocks and scenarios can be used, the most important factors to make the decision are the availability of the feedstock and profitability of the process. Therefore, we decided to perform an economical analysis for the GTE and compare it to the economics of other methods of ethylene production. Data for the commercial methods of ethylene production are available in literature. Similar methodology was used to evaluate the cost of the different techniques of ethylene production. We also analyzed the factors that affect the overall cost of ethylene such as fixed capital cost, utilities, and raw materials. To meet our objective, a flow diagram of GTE proposed by researchers at Texas A&M University was modified, completed, and run using ProMax. In addition, Capcost software was used to estimate the total fixed cost and utilities of the GTE process. Finally, the cost of ethylene production by other methods was compared to the ethylene cost of the GTE process by using a cost index.

2. LITERATURE REVIEW

Most studies in the literature related to ethylene production have been conducted to improve the current process technology. Commercially ethylene is obtained by (1) thermal cracking of hydrocarbons such as ethane, propane, butane, naphtha, kerosene, gas oil, crude oil, etc, (2) autothermic cracking (partial oxidation) of the above hydrocarbons, (3) recovery from refinery off-gas, (4) recovery from coke-oven gas, and (5) catalytic dehydration of ethyl alcohol or ethyl ether.[1] Occasionally, raffinates from aromatics extraction facilities are used as a supplementary raw material. Of the five methods above, small quantities of ethylene are recovered from coke oven gas and from gases produced from crude oil directly [11, 12, 13], but this route to ethylene has, for a variety of technical and economic reasons, so far not gained commercial significance.

In the United States, ethane and propane have been traditional feedstocks for the production of ethylene by the fired tubular heater. In Europe and Japan, where natural gas is not abundant, thermal cracking of naphtha by the fired tubular heater is the most common process. In addition ethylene can be produced by crude oil using an autothermal cracker, and by synthesis gas using Fischer-Tropsch technology. More recently, at Texas A&M University, a new technology was proposed to produce ethylene directly from natural gas.

In this section, I discuss the basic chemistry involved in ethylene production, and I review the general processes and units used to produce ethylene. Then, I present the different processes of ethylene production based upon the most common feedstocks. Of them, ethane, propane, and naphtha, using fired tubular heater are reviewed first. I also introduce ethylene production from crude oil, using autothermic cracking, and synthesis gas from carbon monoxide and hydrogen, using the Fisher-Tropsch process. Finally, I discuss the latest technology of ethylene production directly from natural gas GTE.

2.1 Chemistry of the Ethylene Process

Ethylene, because of its double bond, is a highly reactive compound, which is converted to a multi-intermediates and end end-products on a large scale industrially. The thermal cracking process is the most interesting process to produce ethylene commercially; therefore, I have decided to discuss thermal cracking in more detail than other methods of ethylene production. Even though many types of thermal cracking reactors have been proposed, the fundamental chemistry, mechanism, and kinetics of the process should be the same and independent of the type of the reactor.

In general the starting martial for ethylene production by thermal cracking can be any kind of hydrocarbon. In reality, the choice of starting material is narrowed by economical considerations. The thermal cracking process is fundamentally a dehydrogenation process, accompanied to some extent by polymerization and reactions among products to form the ring structure of aromatics and naphthalenes. As the molecular weight of the feedstock increases, the product complexity increases. Methane can be pyrolyzed to yield mainly ethylene. The reaction is first order taking place at 1000 °C after a protracted induction period, followed by a rapid decomposition and subsequently slower reaction.[14]

Methane, which is an important and abundant fuel, has not been an attractive staring material for ethylene production, because it is thermally stable and has no carbon-carbon bonds. The carbon-hydrogen bond requires more energy to break than the carbon-carbon bond. The C-H bond energy is 93.3 Kcal, whereas C-C energy bond is 71.0 Kcal.[15] The net reaction for methane dehydrogenation is

$$2CH_4 \rightarrow C_2H_4 + 2H_2 \tag{1}$$

The net dehydrogenation reaction of ethane is

$$C_2H_6 \rightarrow C_2H_4 + H_2 \tag{2}$$

In the dehydrogenation of propane four initial reaction steps are conceivable when producing ethylene and propylene; however, according to Sherwood [16,17] and Martin [18] the first two reactions are primary. The reactions are

$$C_3H_8 \rightarrow C_2H_4 + CH_4 \tag{3}$$

$$C_3H_8 \rightarrow C_2H_6 + H_2 \tag{4}$$

$$2C_3H_8 \rightarrow C_2H_8 + 2CH_4 \tag{5}$$

$$2C_3H_8 \rightarrow C_2H_6 + C_3H_6 + CH_4 \tag{6}$$

Because many reactions occur during thermal cracking, it is complicated to determine the rate of the cracking and predict the distribution of the products. Yet, investigations have confirmed that the primary reaction, which splits the original hydrocarbon, is unimolecular and that conversion rates follow the first order kinetics for a wide range of molecular weight and up to high conversion of the original reactant, if there is no distinct equilibrium barrier.[7]

2.2 General Processes for Ethylene Production

In selecting a process for ethylene production, the most important factor is the hydrocarbon feedstock. Although this is controlled by conditions like quantity, quality, and economics studies have shown that as the molecular weight of the feed hydrocarbon increases, ethylenes yield decreases. Generally speaking, the ethylene processes consists basically of four sections: thermal cracking, including quenching, gas compression and gas treatment, ethylene purification, and refrigeration. In the following sections, I discuss each step of ethylene production.

2.2.1 Thermal Cracking Section

The first section of ethylene production process is thermal cracking. Thermal cracking is the heart of an ethylene plant. This section produces all the products of the plant, while other sections serve to separate and purify the products. Additionally, this section has the greatest effect upon the economics of the process. Various types of pyrolysis reactors have been proposed and commercialized for the thermal cracker. These pyrolsis reactors include (1) direct heating (2) indirect heating (3) autothermic cracking and (4) others.

The direct heating process using fired tubular heater is the most common in cracker in an ethylene plant. In this process a variety of the hydrocarbon feedstocks can be used ranging from ethane to gas oil. The fired tubular heater is similar in appearance to the oil or gas furnace that is common in petroleum refinery operation. However, the heat gradient of fired tubular heater is fundamentally different than those of oil or gas furnaces. Another difference is that the cracking temperature for the production of lower olefins is much higher in the fired tubular heater than that used in refinery operations.

Steam is added to the hydrocarbon feed for several reasons: (1) reduce the partial pressure of hydrocarbon, (2) lower the residence time of the hydrocarbon, and (3) decrease the rate of coke formation within the tubes by reaction of steam with carbon to form carbon monoxide and hydrogen.

Thermal cracking of hydrocarbons by indirect heating include the pebble bed reactors, the fluidized bed reactors, and regenerative furnace. These reactors release much more energy in the reaction space than do tubular furnaces of the same volume. Also, they have fewer materials of construction problems at high temperature than tubular heater. Therefore, shorter residence times and higher pyrolysis temperatures can be employed in the indirect heating processes than in the fixed tubular heater. Even though construction costs seem to be more expensive and the operation to be more complex, ethylene yield by indirect heating is higher than that obtained in the fired tubular heaters. One of the advantages of the indirect heating reactors is that crude oil and heavy fuel oil can be used as feedstocks, because the coke byproduct can be removed continuously or intermittently in the process.[7]

Many autotherma processes have been proposed, and most of them, except the Badische Anilin- und Soda-Fabrik (BASF) crude oil process using a fluidized bed of coke, produce acetylene as a main product and ethylene as a by-produce or as a co-product. Therefore, these autothermic processes are advantageous when both acetylene and ethylene are desired products and are disadvantageous when only ethylene and propylene are desired, because acetylene removal is costly. In this process the endothermic heat of cracking is balanced by the heat of combustion of a portion of the hydrocarbon feed with air or oxygen so that the overall reaction is self-sustaining. When combustion gases are used in the process as a heat carrier to crack the hydrocarbon, the process is grouped with the autothermic cracking for the reason that they use air or oxygen for combustion of the fuel, and the combustion gases are mixed with the cracked products. In the Autothermal process atmospheric pressure is used; and hydrocarbon feedstocks, air, or oxygen and fuels are separately preheated to about 1100°F to increase the yield and reduce oxygen consumption.

Among other processes are catalytic cracking, molten metal bath, and a process synthesizing ethylene from carbon monoxide and hydrogen. The catalytic effect to produce ethylene and propylene by thermal cracking of hydrocarbons is not so important; especially above 1300 °F. A catalyst process is sometimes preferable for several reasons: it reduces the cracking temperature or avoids coke formation, or increases the selectivity of specific olefins or aromatics. The synthesis of ethylene by carbon monoxide and hydrogen may be interesting as a method to ethylene from coal and heavy oils. The selectivity of ethylene is high (about 95%), but carbon monoxide conversion is quite low (10-35%). With such a low conversion, it may be that ethylene recovery by solvent absorption would require an appreciably lower capital investment than recovery by low temperature fractionation. However, the high cost of synthesis gas appears to be a fundamental obstacle to commercialization of this process.[7]

The pyrolsis gas leaving the cracker usually has a temperature in the range of 375°C to 500°C in the case of naphtha pyrolsis and typically from 500°C to 600°C in the case of gas oil pyrolsis. The outlet temperature depends upon the amount of the carbon deposits in the transfer line exchanger.[1] Quenching of the conversion product or rapid temperature reduction is important to prevent the decrease of ethylene yields caused by secondary reactions. This is carried out either by transfer line exchangers or by injecting water and oil.

2.2.2 Gas Compression and Treatment Section

In addition to the thermal cracking section, the sections for removal of acid gases, drying of the cracked gases, removal of acetylenic compounds, and purification of ethylene are also very important, because an efficient ethylene plant is the result of the integration of theses process sections and because, in respect to cost, the thermal cracking section is the only about 20-30% of the whole plant. In addition, the goal is to produce ethylene with high purity above 99.9%.

Most ethylene processes call for compression of the pryolsis gas leaving the quench tower. This compression is important for subsequent cryogenic treatment. Consequently, the cooled cracked gas leaving the water tower is compressed in four to five stages. Plants based upon gaseous feedstock generally employ four stages, while many naphtha-and gas oil-based plants employ five stages of pyrolysis gas compression. Between compression stages, the cracked gas is usually cooled in water-cooled exchangers. Water and hydrocarbons condensed between stages are separated from the pyrolysis gas in inter-stage separators.

Hydrogen sulfide and carbon dioxide are removed from pyrolysis gas between the third stage and fourth stage of the compression system. This location is optimum because the actual gas volume has been reduced significantly in the first three stages of compression while acidic components are still present in the gas stream.[1] Acid gas produced in thermal cracking must be removed before the first major fractionation step. In removing acid gases such as carbon dioxide and hydrogen sulfide, non-regenerative caustic washing followed by water washing is employed in the most of the existing plants and proves to be most economic except for autothermic cracking processes in which the acid gas content is high. In this case, a combination of conventional amine absorption followed by caustic washing seems to be common. The pyrolysis gas leaving the caustic scrubber contains less than 1 ppm of acid gases and hence assures that the final products of the plant will meet specification in this respect.

Compressed cracked gas usually is dried to reduce the moisture content to 1 ppm or less and avoid problems with freezing and hydrate formation in downstream low temperature equipment. In drying the cracked gases, alumina, silica gel, and molecular sieves are used commercially. Among them, molecular sieves seem to have an economic advantage over conventional desiccants because of their higher desiccant activities and lower regeneration temperatures.[7] Silica gel is not desirable where large quantities of propylene are present because it tends to promote polymerization and resultant loss of temperature control.

Removal of acetylene and acetylenic materials from the process gas is very important in manufacturing polymer-grade ethylene. There are basically two methods to remove acetylene form cracked gases. One is selective hydrogenation over nickel, copper, cobalt, iron, palladium, and several other metallic oxides. The other is selective absorption by acetone, dimethyl formamide (DMF), and others.[7] Acetylene removal by selective hydrogenation may be accomplished at several points in the recovery and purification stages: before acid gas removal, between acid gas removal and drying section, between drying section and demethanizer, in the ethane-ethylene stream, and in the final ethylene stream.

2.2.3 Recovery and Purification Section

After the cracked gases have been quenched, compressed, freed of the acid gases, dried, and after the removal, in some cases, of acetylenic materials, they generally contain hydrogen and light hydrocarbons in the C_1-C_6 range. Depending upon the cracking method employed, carbon monoxide and nitrogen also may be present. Low temperature straight fractionation, absorption, and selective adsorption are three different methods to recover and purify ethylene. At this stage, different scenarios can be used to produce ethylene. The aim of this section is to separate ethylene from hydrogen and methane fractions, ethane and propane fractions, and heavier hydrocarbons. Within the separation process, there are several possible schemes. There are also some temperature limitations; there are maximum temperature limits to prevent polymerization fouling, and minimum temperature limits to prevent hydrocarbon freezing or hydrate formation. In addition, commercial separation processes operate at four ranges of pressure: 450-600 psia, 100-150 psia, 70-90 psia, and 30-40 psia.[1] The most popular is the 450-600 psia because it offers an attractive combination of purity, recovery, efficiency, and investment for large ethylene plants.

In ethylene purification section, demethanized process streams are introduced to the de-ethanizer in most cases. The de-ethanizer is a simple tower refrigerated by propane or propylene to make reflux. The net overhead from the de-ethanizer flows to an ethylene-ethane separator. This is the second most costly separation step in an ethylene pant because the volatility is low and a large amount of reflux is required.

2.2.4 The Refrigeration Section

The separation of pyrolysis gas through condensation and fractionation at cryogenic temperatures requires external refrigeration and is an important part of the ethylene system. Today, propylene-ethylene cascade refrigeration is the most commonly used in ethylene plants, because it is suitable and readily available in the plant. An ethylene refrigerator has two or three stages for a total of between five and seven stages for the entire refrigeration cascade. Reflux ratios in the columns are selected carefully to avoid large refrigeration consumption.[6]

In the next section, I discuss various processes of ethylene production. These processes use the most common technology and feedstock of ethylene production including the latest technology of ethylene production directly from natural gas (GTE).

2.3 Fired Tubular Heater

Typical equipment for direct heating is the fired tubular heater. Ethane, propane, and light naphtha feedstocks use fired tubular heaters for ethylene production. In this part, I discuss the three common feedstocks that employ fired tubular heating for the ethylene production. All the data and the flow sheet are based upon a private report from the process economics program at Stanford Research Institute (SRI) about ethylene processes and economics. For comprehensive analysis of each process, you can refer to SRI report for ethylene. It presents the process in more detail and provides more information about each unit and stream in the process.

2.3.1 Ethane

Ethane is obtained on industrial scale from natural gases and as a byproduct of petroleum refining. After methane, ethane has the second highest composition in natural gas. Ethane is separated most efficiently from methane by liquefying it at cryogenic temperatures. Various refrigeration strategies exist, but the most economical process presently in wide use employs turbo-expansion, and can recover over 90% of the ethane in natural gas.[19] The principal use of ethane is in chemical industry, mainly, in the production of ethylene by steam cracking. Ethane is favored for ethylene production because the steam cracking of ethane is fairly selective for ethylene. Ethane may be cracked alone or as a mixture with propane is shown in figures 3 and 4.

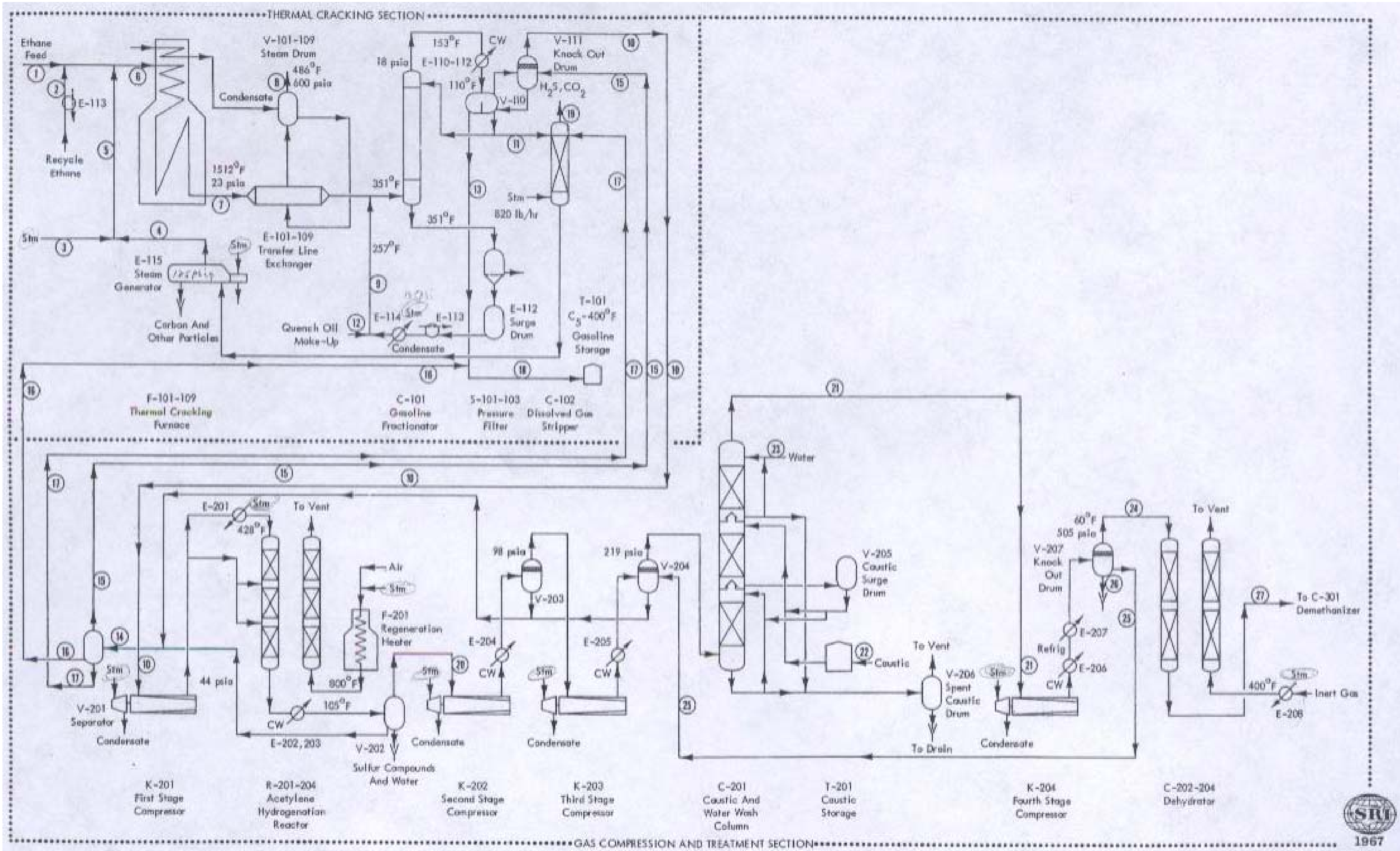

Figure 3: Ethane pyrolysis by fired tubular heater (thermal cracking and gas compression and treatment sections)

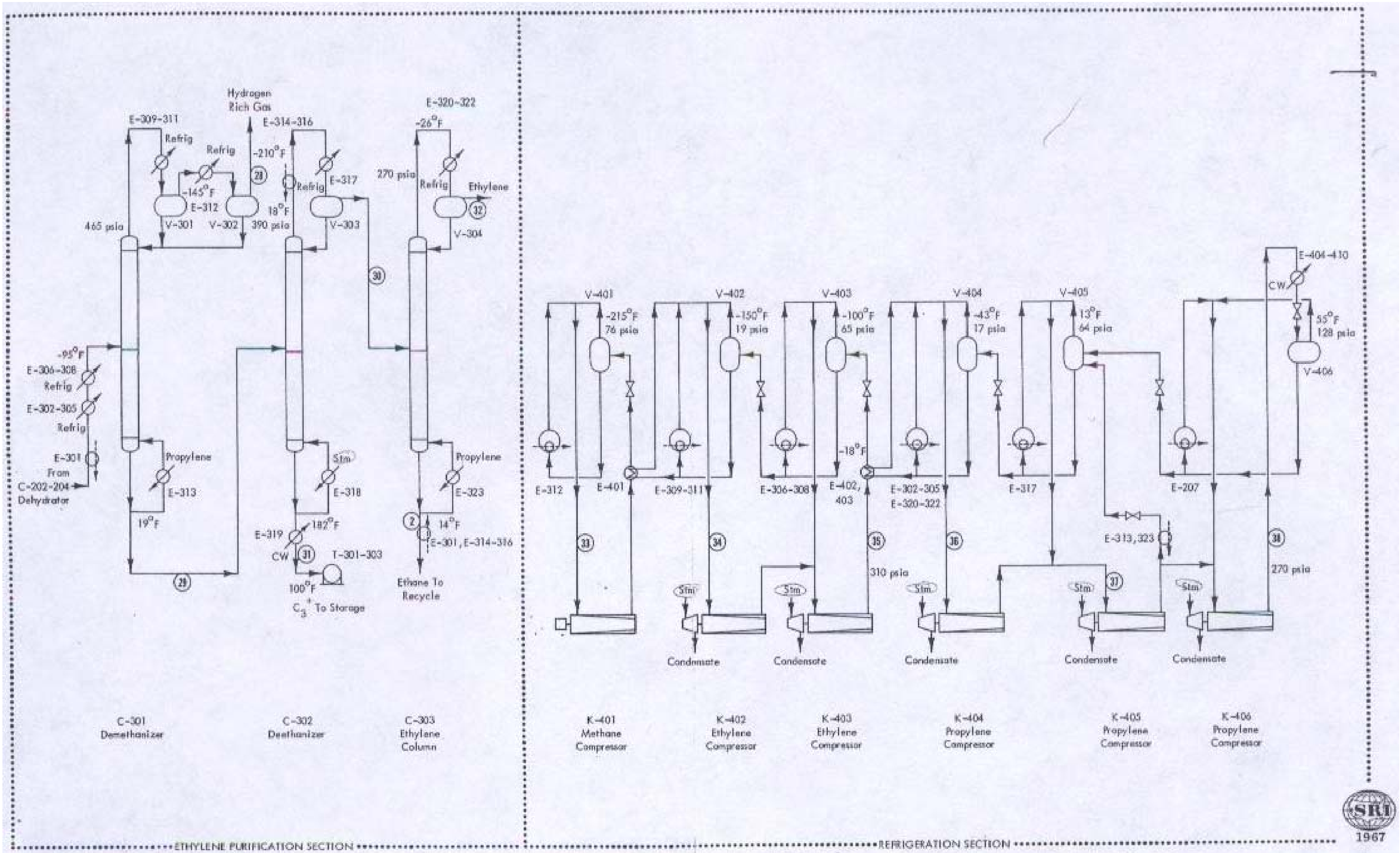

Figure 4: Ethane pyrolysis by fired tubular heater (ethylene purification and refrigeration sections)

The fresh feed of ethane is combined with the recycle from the ethylene column and charged to the thermal cracking furnaces. Dilution steam is added to ethane before it enters the cracking furnaces to reduce the partial pressure of ethane and lower the residence time of ethane in the high temperature zone, which decrease the rate of coke formation within the tubes. The mixture of ethane and steam is preheated in the convection section of the furnace, and the ethane cracks in vertical tubes. The cracked gas leaves the furnace at 1512 °F and 32 psia and is quickly quenched to 500 °F in the transfer line exchangers. It generates 600 psia steam at the rate of 2.46 pounds per pound of ethylene produced. The dilution stem is a closed loop. The dilution steam is condensed in the gasoline fractionator overhead condensers and the condensates are separated in the reflux drum. Then the gas goes to the dissolved gas stripper, where the dissolved gases such as acetylene, ethylene, propylene, hydrogen sulfide, and carbon dioxide are stripped from of the dilution steam condensates by the stripping steam. This permits their use for generating low pressure steam in the steam generators. If C_5 cracked gasoline contains a significant amount of dissolved C_4 and lighter components, these lights ends are stripped and recovered as an overhead vapor from the lighter ends stripper. Then, the cracked gasoline is stored in gasoline storage tanks. The quench oil absorbs most byproducts from the previous stages, and they are separated from quench oil by a pressure leaf-filter. Because a small amount of quench oil is lost with the byproducts, the fresh quench oil is made up to keep the viscosity and the slurry content of the quench oil low. The cracked gas of C_4 and lighter components are separated in the gasoline fractionator reflux drum and fed to the first stage compressor. The cracked gases are compressed in four stages.

Between the compression stages hydrogenation reactors hydrogenate acetylene compounds. Organic sulfide compounds are converted into hydrogen sulfide and mercaptans that are removed by caustic washing. Acid gases are also scrubbed between compression stages by caustic soda and washing water. The effluent gas leaving the compression section is then cooled by water and chilled by high-level propylene refrigerant. Molecular sieve dryers dry the chilled gas and pass it to the low temperature demethanizer. Before the gas enters the demethanizer, it is cooled to –95°F by successive stages of heat exchangers. The demethanizer is reboiled with condensing propylene refrigerant and the overhead vapor is partially condensed by low-pressure ethylene refrigerant. The net bottom stream of the demethanizer is charged to the de-ethanizer. The overhead vapor from the de-ethanizer is partially condensed by heat exchange and propylene refrigerant. The bottom stream leaving the de-ethanizer contains C_3 and C_4 hydrocarbons that are stored in C_3+ storage tanks. The net overhead from the de-ethanizer is the ethylene-ethane. This stream is then fed to the ethylene column to separate ethylene.

The overhead vapor of the ethylene column is partially condensed and high purity polymer grade ethylene is recovered as a vapor product. The refrigerant system provides the bulk chilling and condensing duties for the plant.

2.3.2 Propane

Propane C_3H_8 is normally a gas, but it is compressible to a liquid that is transportable. It is derived from other petroleum products during oil or natural gas processing. Propane, also known as liquefied petroleum gas (LPG), can be mixture of propane with small amounts of propylene, butane and butylenes. Propane is a byproduct of natural gas and petroleum refining. Propane is used as a feedstock for ethylene production. The production of ethylene from propane is similar to the process of ethylene production from ethane. Ethylene production from propane consists of four sections: thermal cracking, gas compression, gas treatment and ethylene purification, and refrigeration. The flow sheet of the process is figures 5, 6, and 7.

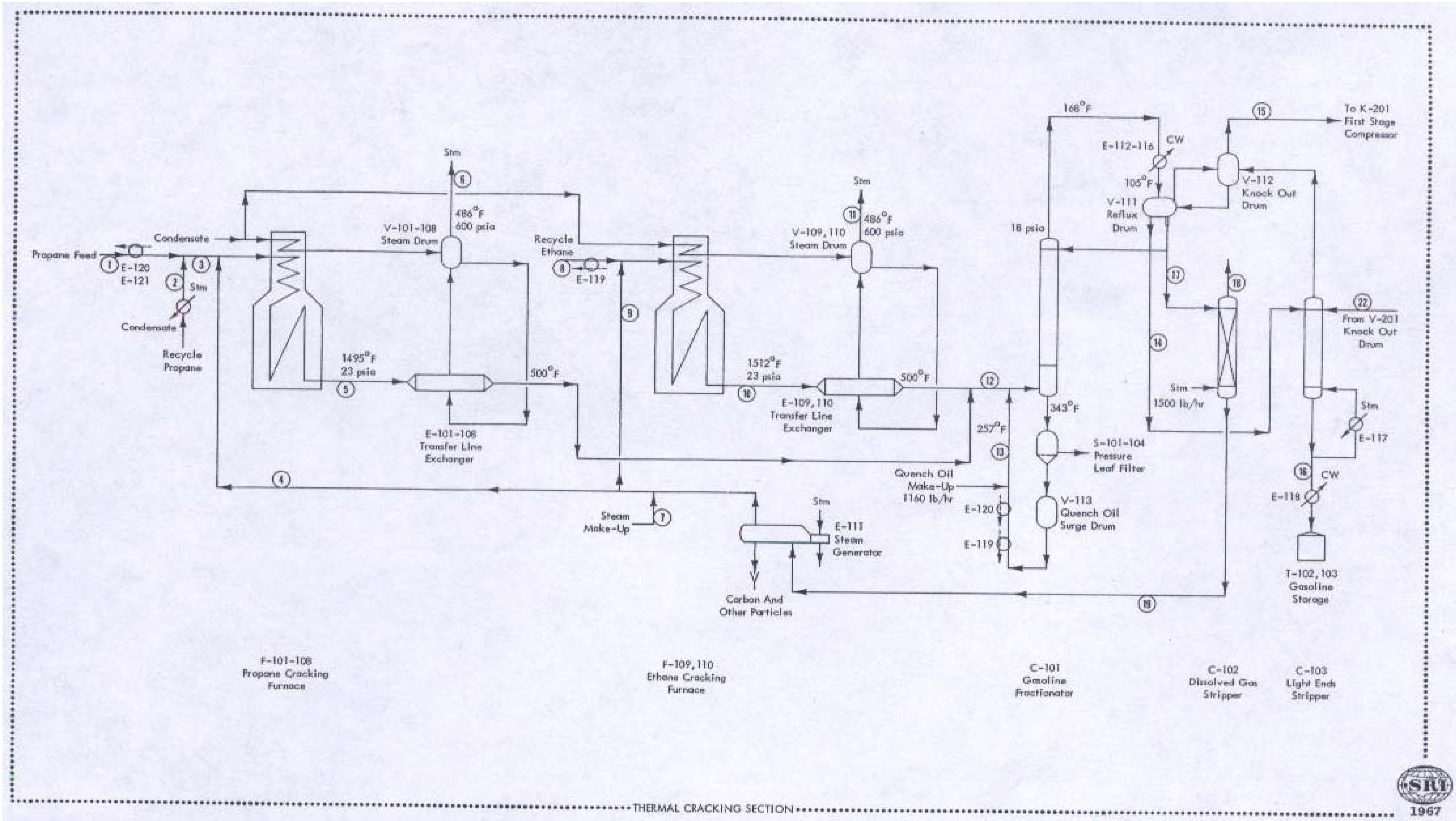

Figure 5: Propane pyrolysis by fired tubular heater (thermal cracking section)

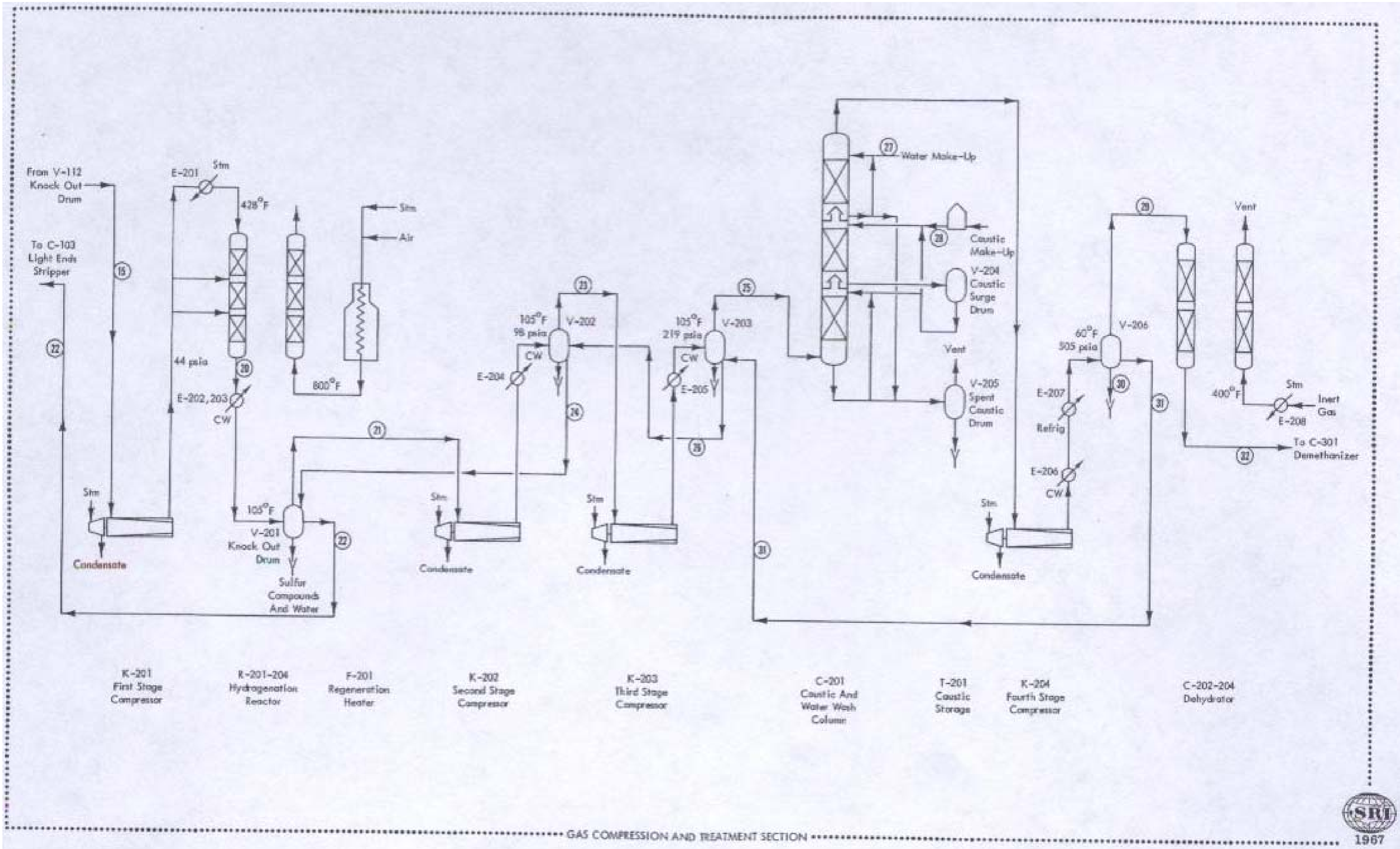

Figure 6: Propane pyrolysis by fired tubular heater (gas compression and treatment sections).

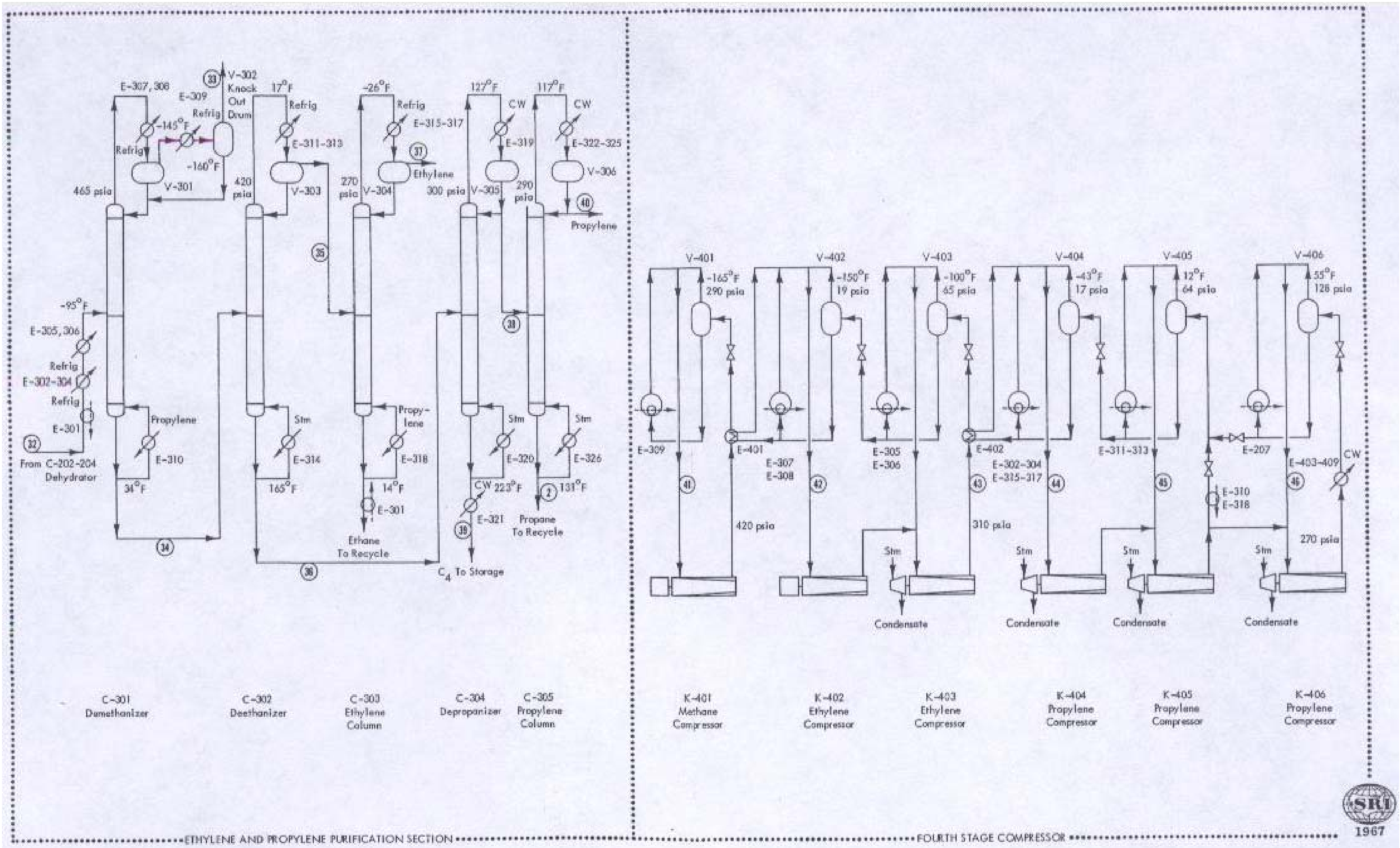

Figure 7: Propane pyrolysis by fired tubular heater (ethylene and propylene purification and fourth stage sections).

Fresh feed of propane is combined with recycled propane from the propylene column and charged to the propane-cracking furnace. To increase the overall yield of ethylene from fresh propane, ethane is co-produced at the rate of 6 pounds per pound of propane feed. Dilution steam is added to the propane before it enters the cracking furnace to reduce the partial pressure of propane and lower the residence time of the propane in the high temperature zone, thus decreasing the rate of coke formation in the tubes.

Similar to ethane furnaces, the mixture of propane and steam is preheated in the convention section of the furnace, and propane cracks in vertical tubes. The cracked gas exits the furnaces at 1495 ºF and 23 psia, and is quenched to 500ºF in the transfer line exchanger where it generates 600 psia steam at the rate of 4 pounds per pound of ethylene produced. The exiting stream from transfer line exchangers is further quenched with circulating quench oil and fed to the gasoline fractionator. The dilution steam is condensed in the gasoline fractionator overhead and the condensates are separated in the reflux drum. Then, it is sent to the dissolved gas stripper where the dissolved gases such as acetylene, ethylene, propylene, hydrogen sulfide, and carbon dioxide are stripped out of the diluted steam condensates by the stripping steam thus permitting their use for generation of low pressure steam in the steam generator. If the cracked gasoline contains an appreciable amount of dissolved C_4 and lighter hydrocarbons, these lighter ends are stripped and recovered as an overhead vapor form the light ends stripper. The cracked gasoline is stored in gasoline storage tank. Carbon dust, tar, and polymerized higher hydrocarbons are byproducts produced in small quantities and are mostly absorbed in the quench oil. Pressure leaf filters separate these components from the quench oil and fresh quench oil is added to the process to make up for the quench oil lost in removing carbon dust and tars and to keep the viscosity and the slurry content of the quench oil low. After that the cracked gases of C_4 and lighter components are separated in the gasoline fractionator reflux and fed to the next section of the process, which is compression.

Similar to the compression section of ethane process, the cracked gases are compressed and acetylenic compounds are hydrogenated and removed between the compression stages. Acid gases are also scrubbed and washed out between compression stages. Then, the gas leaving the compression section is chilled by high-level propylene refrigerant.

Before sending the gas to the demethanizer, it is dried by molecular sieves dryers. The demethanizer overhead vapor is partially condensed by low-pressure ethylene refrigerant, while it is reboiled with condensing propylene refrigerant. The net bottom of the demethanizer is charged to the de-ethanizer. The overhead of the de-ethanizer is condensed by propylene refrigerant, and a C_3-C_4 stream as a bottom product feeds to the de-propanizer. A propane-propylene mixture is separated as the overhead product, and a C_4 bottom stream is stored. Cooling water condenses the overhead vapor from the de-propanizer, and the net overhead liquid is charged to the propylene column to recover polymer grade propylene.

Propane is obtained as a bottom product and is recycled to the propane-cracking furnace. The high purity ethylene product is recovered from the net overhead vapor of the de-ethanizer by ethylene column.

2.3.3 Naphtha

Naphtha, an important feedstock for ethylene production, is a collective of liquid hydrocarbon intermediate oil refining products. It is a mixture of hydrocarbons in the boiling point range of 30-200 ºC. For the naphtha cracker process, typical feedstocks are light naphthas (boiling range of 30-90 ºC), full range naphthas (30-200 ºC), and special cuts (C6-C8 raffinates).[5] Naphtha is obtained in petroleum refineries as one of the intermediate products from the distillation of crude oil. As mentioned earlier, thermal cracking by fired tubular heaters using naphtha as a feedstock is more common in Europe and Japan. The processing of light naphtha to ethylene is similar to the ethane and propane processes. The flow diagram of the process is figures 8, 9, and10.

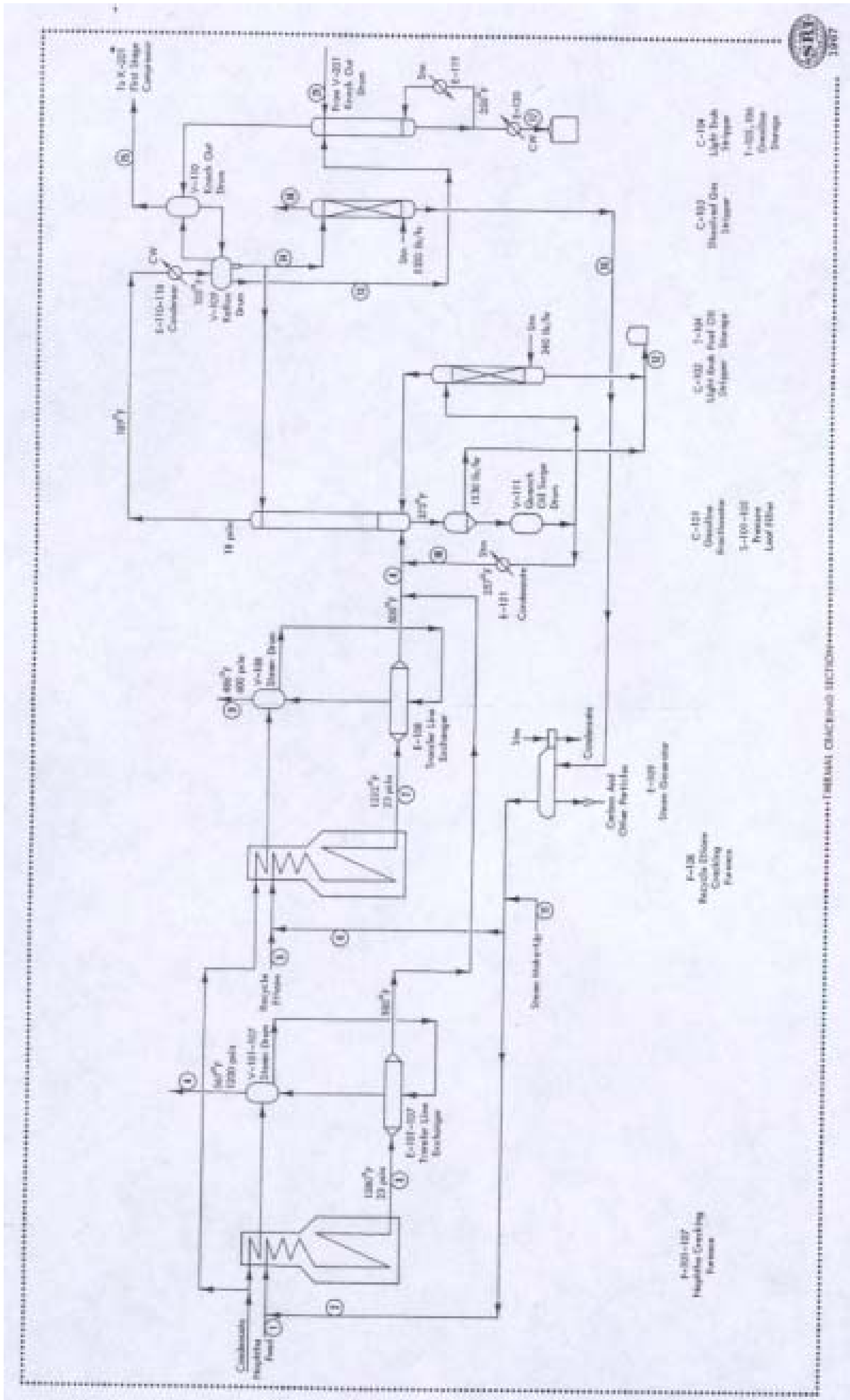

Figure 8: Naphtha pyrolysis by fired tubular heater (thermal cracking section)

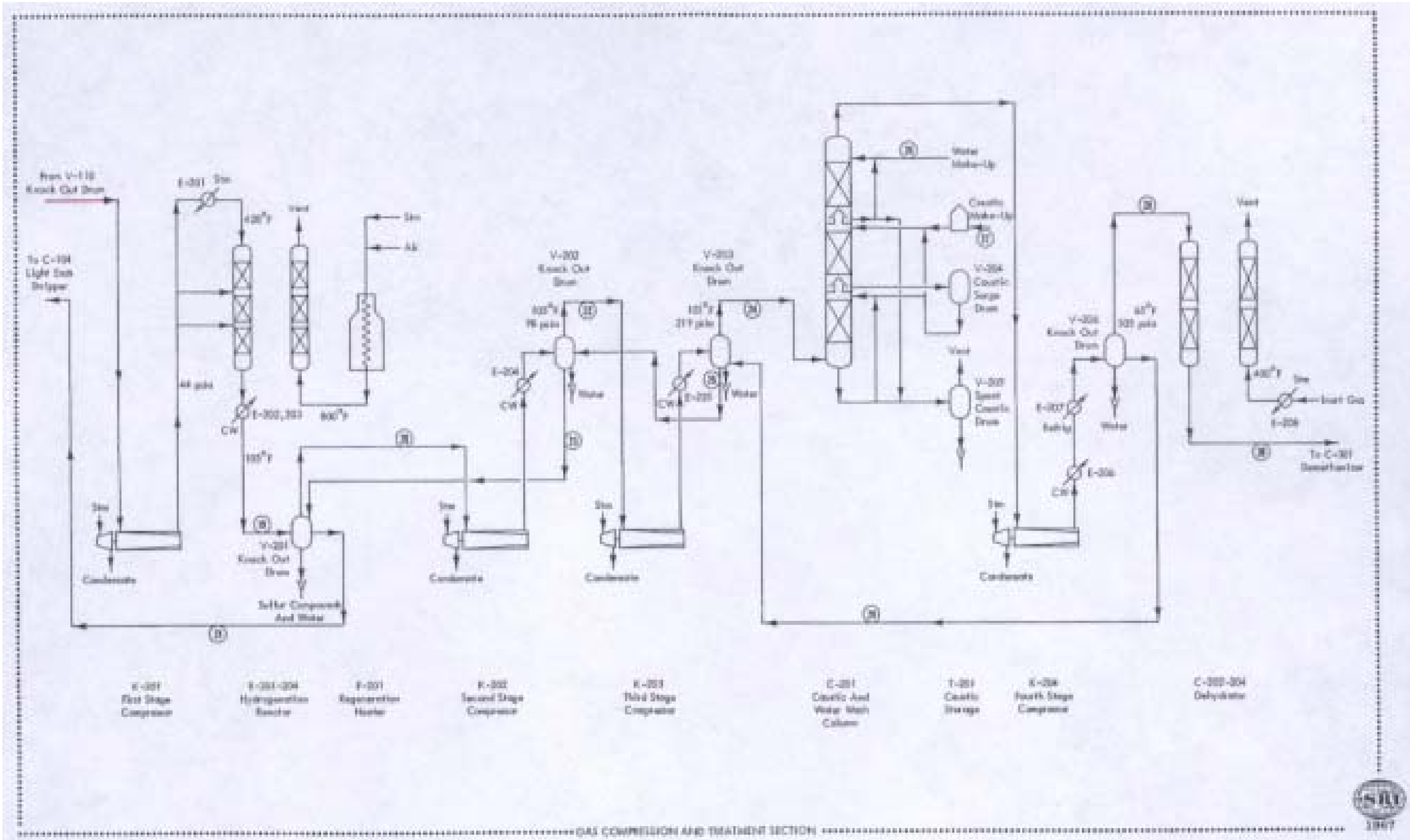

Figure 9: Naphtha pyrolysis by fired tubular heater (gas compression and treatment section).

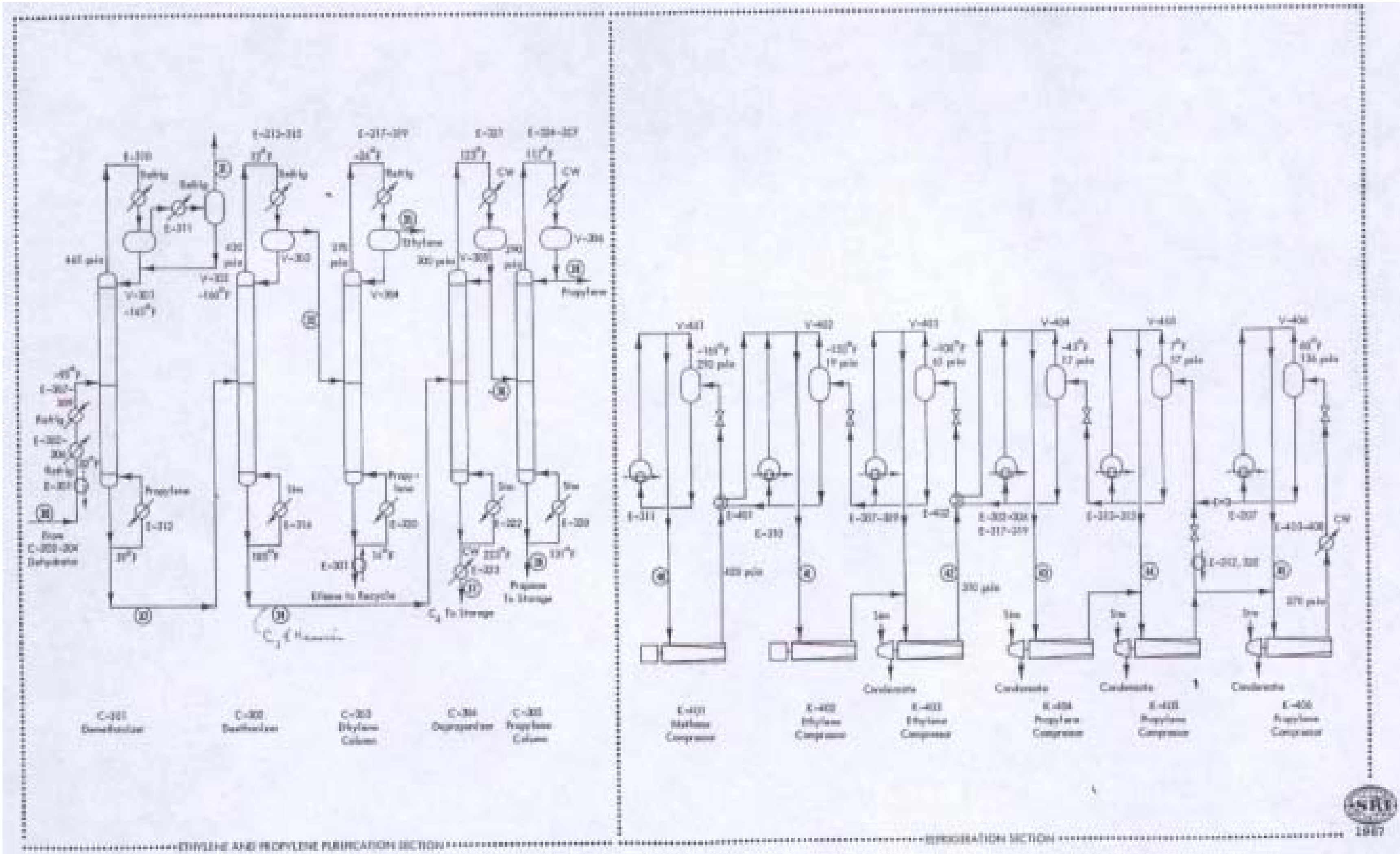

Figure 10: Naphtha pyrolysis by fired tubular heater (ethylene and propylene purification and refrigeration sections).

Light naphtha having the following specification:

ASTM boiling range	90-260 F
Paraffin content	86 wt%
Naphthene content	11wt%
Aromatic content	3wt%

is fed to the naphtha cracking furnace.[1] By-product ethane, which is obtained in the ratio of 4.5 pounds per pound of naphtha feed and separated as the bottom product of the ethylene column, is cracked at different conditions in the ethane cracking furnace to increase the overall yield of ethylene. Like previous processes, diluted gas is added to the hydrocarbon feedstock before it enters the furnace to decrease the partial pressure of naphtha, minimize coke formation, and minimize residence time in the furnace. The mixture of naphtha and steam is preheated before it is charged into the cracking furnaces. The cracked gas from light naphtha leaves the furnace at 1580°F and 23 psia. Then it is quenched to 580°F in the transfer line exchanger and generates 1200 psia steam at the rate of 5.6 pounds per pound of ethylene produced.

Circulating quench oil, a by-product fuel oil, further quenches the stream exiting the transfer line exchangers. Then, it is fed to gasoline fractionaor, where it separates the quench oil as a bottom and the cracked gasoline and lighter components as the overhead products. The dilution steam is condensed in the gasoline fractionator overhead condensers and the condensates are separated in the reflux drum. Then, a dissolved gas stripper removes dissolved gases, such as acetylene, ethylene, propylene, hydrogen sulfide, and carbon dioxide, from the diluted steam condensates using stripping steam.

Dissolved C_4 and lighter components, which condense in the reflux drum of the gasoline fractionator, are stripped and recovered as an overhead vapor from the light ends stripper. Then, the cracked gasolines are stored. Steam in the stripper separates carbon dust and polymerized higher hydrocarbons that are absorbed in the quench oil. At the same time, a small amount of fresh oil is added. Light ends-free oil is stored in the fuel oil storage tank. After the cracked gas of C4-and-lighter components are separated in the gasoline fractionator reflux, it is fed in to compression section, where the cracked gas is compressed and acetylenic compounds are removed between compression stages. In addition, acid gases are also scrubbed and removed using the same technique used in the ethane and propane processes. Then, the exiting gas from the compression section is cooled by water and chilled by high-level propylene refrigerant.

The chilled gas from the final stage knockout drum is dried in the molecular sieve dryers before passing to the demethanizer. The net bottom steam from the demethanizer is charged to the de-ethanizer. Propylene refrigerant is used to condense the overhead vapor of the de-ethanizer. The bottom product of the de-ethanizer is fed to the de-propanizer that further separates propane-propylene from C_4 components, which are stored. The net overhead vapor from the de-propanizer is condensed and charged to the propane column to recover propylene. Propane, obtained as a bottom product of the propylene column, is stored. The net overhead vapor from the de-ethanizer is the ethylene-ethane stream, and these components are separated in the ethylene column. The overhead vapor is partially condensed and high purity ethylene is recovered as a vapor product.

2.4 Autothermic Cracking in Fluidized Bed

Most of the Autothermic cracking processes produce acetylene as a main product and ethylene as a by-product. Most of these processes operate at atmospheric pressure and hydrocarbon feedstock, air, or oxygen and fuels are preheated to about 1100 ºF to reduce the oxygen consumption and increase the yield. Crude oil is one of the most common feedstocks for authothermic cracking in fluidized bed ethylene production.

2.4.1 Crude Oil

The process of ethylene production by autothermic cracking is based upon the thermal cracking of crude oil using fluidized beds. It consists of four sections: thermal cracking, gas compression, and treatment, propylene and ethylene purification, and refrigeration. The process diagram is figures 11, 12, and 13.

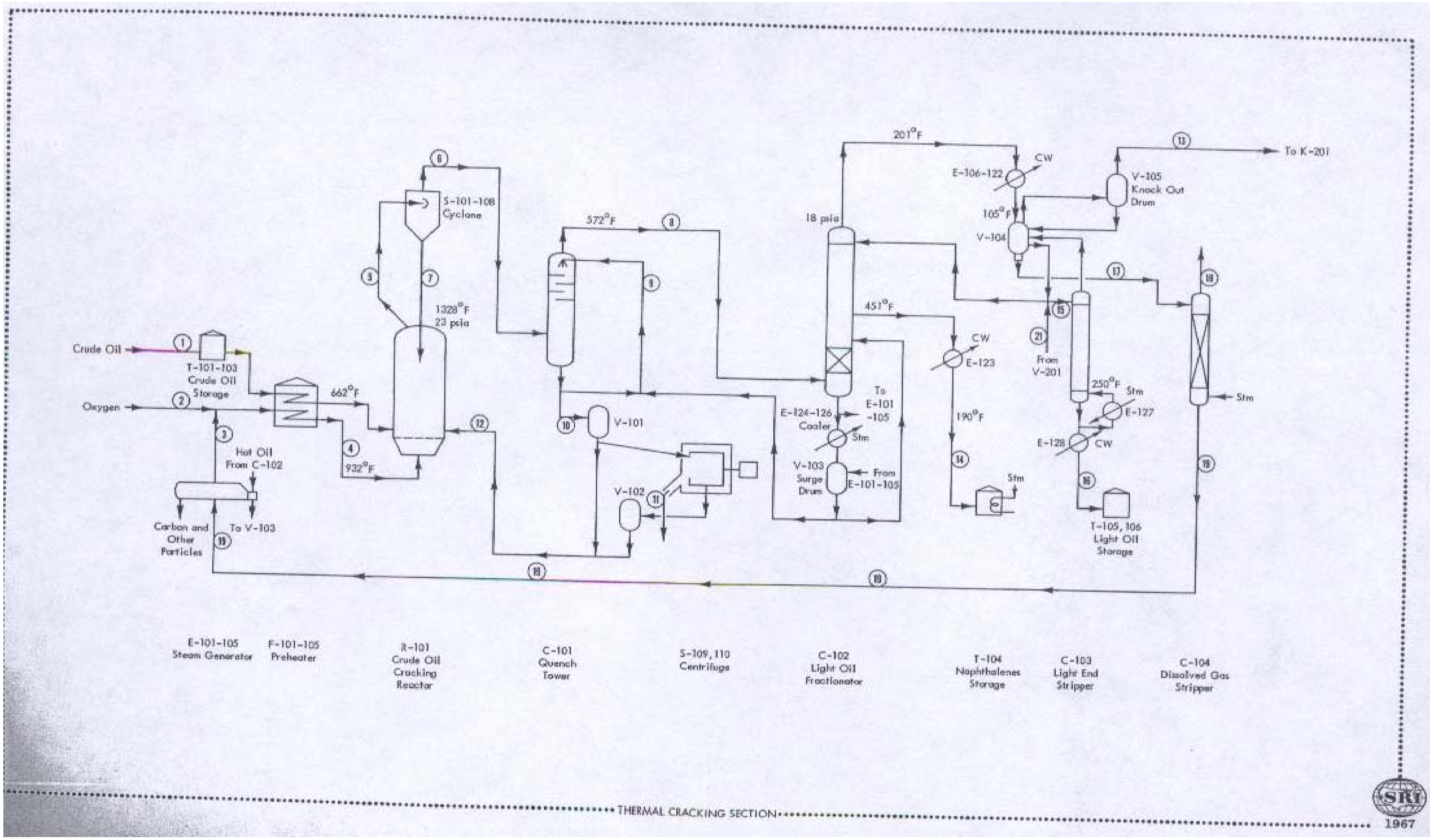

Figure 11: Ethylene from crude oil by autothermal cracking (thermal cracking section).

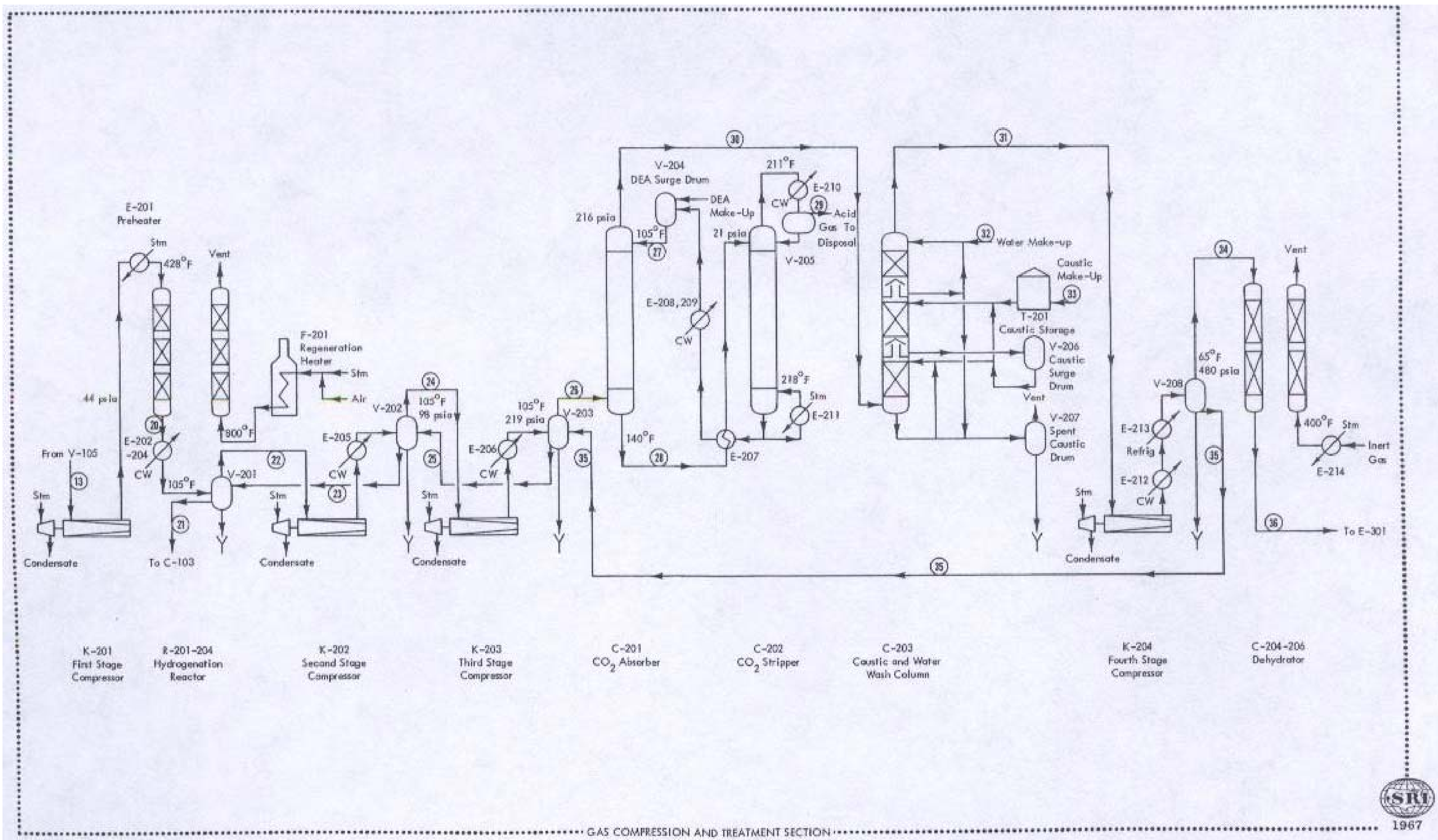

Figure 12: Ethylene from crude oil by autothermal cracking (gas compression and treatment section).

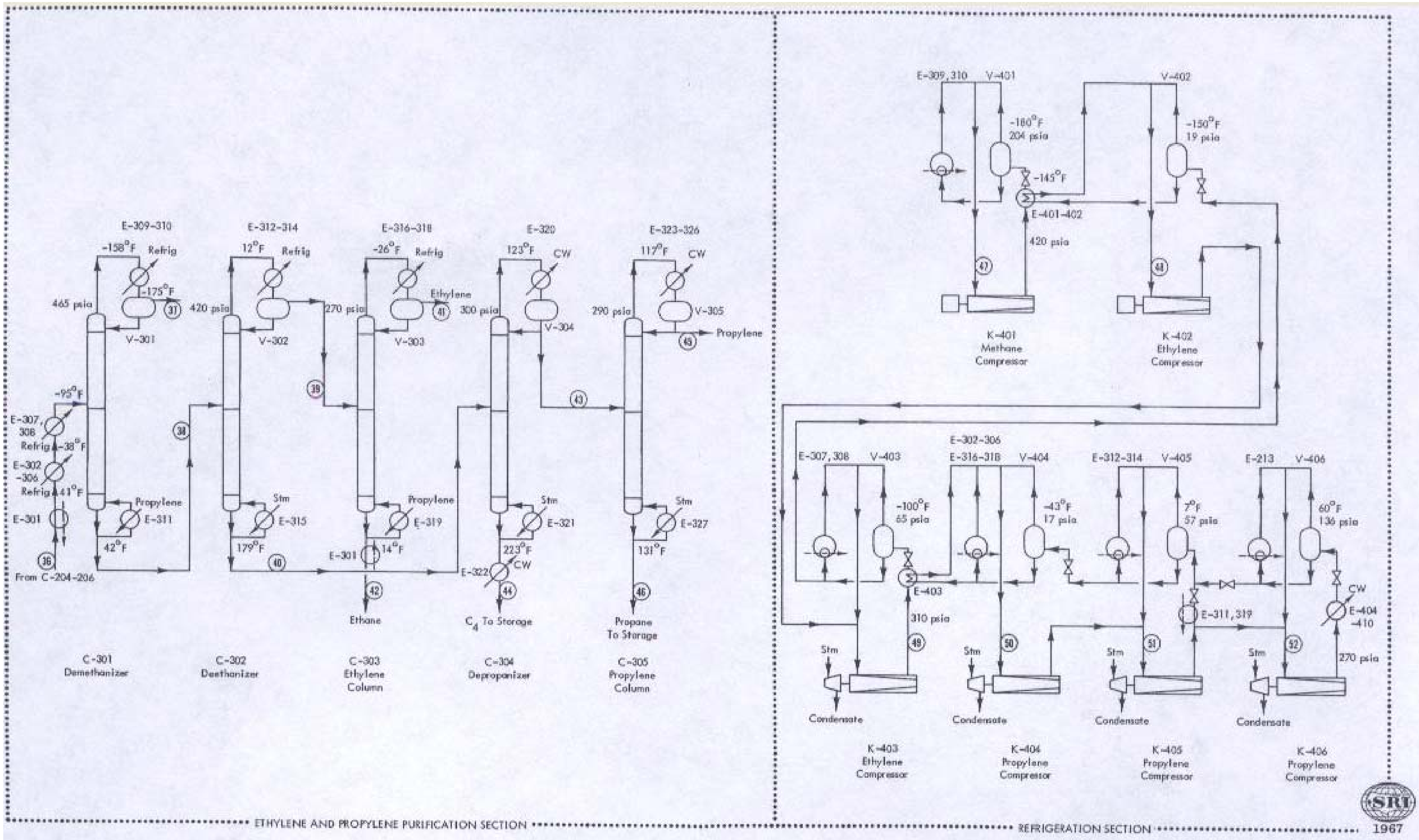

Figure 13: Ethylene from crude oil by autothermal cracking (ethylene and propylene purification and refrigeration ssections)

Crude oil is preheated to 662°F in the furnace and fed to the reactor. Simultaneously, a mixture of steam and oxygen is also preheated to 932 °F and fed to the bottom of the reactor. The exothermic reaction of oxygen and the crude oil forms petroleum coke. This exothermic reaction provides the heat for the endothermic cracking reaction. The high boiling oil is recycled and introduced into the reactor together with the crude oil. The coke formed in the reactor is fluidized using steam, which is passed upward through the reactor. The cracked gas and vapor pass to a cyclone, where most of entrained solids are removed and returned to the fluidized bed by gravity. Then, all gases and solids are fed to the quench tower, where they are cooled by injection of recycled cracked oil. Solids such as the residual coke, soot, and dust are removed from the cracked gases, and the cracked heavy oil carrying these solids discharges into a surge tank. Then, the cracked gases and vapors from the quencher are fed to the bottom of the light oil fractionator together with the vaporized quench oil. From the bottom of the column, the heavy oil is drawn off, and a part of it is used to generate the clean steam in the heavy oil cooler. The rest of heavy oil is charged to the steam generator to generate the process steam from steam condensates with dissolved hydrocarbon and phenolic compounds. Part of the heavy oil is return to the quench and washes the cracked gases and vapors. The rest of heavy oil is recycled to the quench tower. A naphthalene friction is withdrawn from the side steam of the column, cooled and stored. The low boiling oils pass overhead from the column together with the cracked gases and are condensed together with the fluidizing steam. The light steam is separated from the steam condensates in the separator. A major part of it returns to the column as reflux and the rest is stored. Then, the condensates in the separator are sent to the dissolved gas stripper, where steam strips the dissolved gasses such as acetylene, ethylene, propylene, hydrogen sulfide, and carbon dioxide from the condensates thus permitting their use for generation of low pressure fluidizing steam in the steam generator.

The uncondensed gases in the separator are fed to the compression section. Between the compression stages, acetylenic gases are hydrogenated and then removed by caustic washing. Then, carbon dioxide and hydrogen sulfide are removed by absorption in the CO_2 absorber. Gases to be purified pass through the absorber and stripper to remove acid gases. The gases in the CO_2 absorber, contains carbon dioxide that is completely removed by several caustic soda washes followed by water washing. Then, the compressed gases are cooled by water and chilled to condense out as much water as possible before they are passed to the dehydrator. Before the dried gases are fed to the demethanizer, they are chilled by successive stages of heat exchangers.

The demethanizer is reboiled with condensing propylene refrigerant and the overhead vapor is partially condensed using methane refrigerant. The uncondensed gas is a mixture of hydrogen, nitrogen, carbon monoxide, and methane. The net bottom stream from the demethanizer is charged to the de-ethanizer. The C_3-C_4 stream is the bottom product of the de-ethanizer and is charged to the de-propanizer, which separate propylene-propane mixture from the C_4 streams. The net overhead liquid of the de-propanizer is fed to the propylene column to obtain propylene as an overhead product. The net overhead vapor from the de-ethanizer is the ethylene-ethane stream, and ethylene is obtained from the top of the ethylene column. Ethane is the bottom the bottom product of the column.

2.5 Synthesis Gas

The method for synthesizing liquid hydrocarbon from carbon monoxide and hydrogen is widely known as the Fischer-Tropsch process, which is a catalyzed chemical reaction process. Typical catalysts used contain iron and cobalt. The synthesis of ethylene and methane form carbon monoxide and hydrogen is expressed by the following chemical equations:

$$2CO + 4H_2 = C_2H_4 + 2H_2O \tag{7}$$

$$CO + 3H_2 = CH_4 + H_2O \tag{8}$$

The Fischer-Tropsch (FT) process was invented in petroleum poor but coal rich Germany to produce liquid fuels in 1920's, and it also was used in World War II.

As a result, this process for the production of ethylene form carbon monoxide and hydrogen seems to have no direct interest today.

However, FT process has potential interest as a process starting from coal or heavy fuel oils, or for the countries where petroleum sources are not abundant.

2.5.1 Carbon Monoxide and Hydrogen

FT gas to liquid processes or GTL converts natural gas into hydrocarbon fuels via syngas as an intermediate. This process can use gas deposits in locations where it is not economical to build a pipeline. It is predicted that natural gas should be the next major source of energy after the crude oil sources are depleted. The process for the production of ethylene from carbon monoxide and hydrogen contains four sections: ethylene synthesis, gas treatment, the ethylene recovery, and the refrigeration. Figures 14 and 15 are the FT process to synthesize natural gas to ethylene.

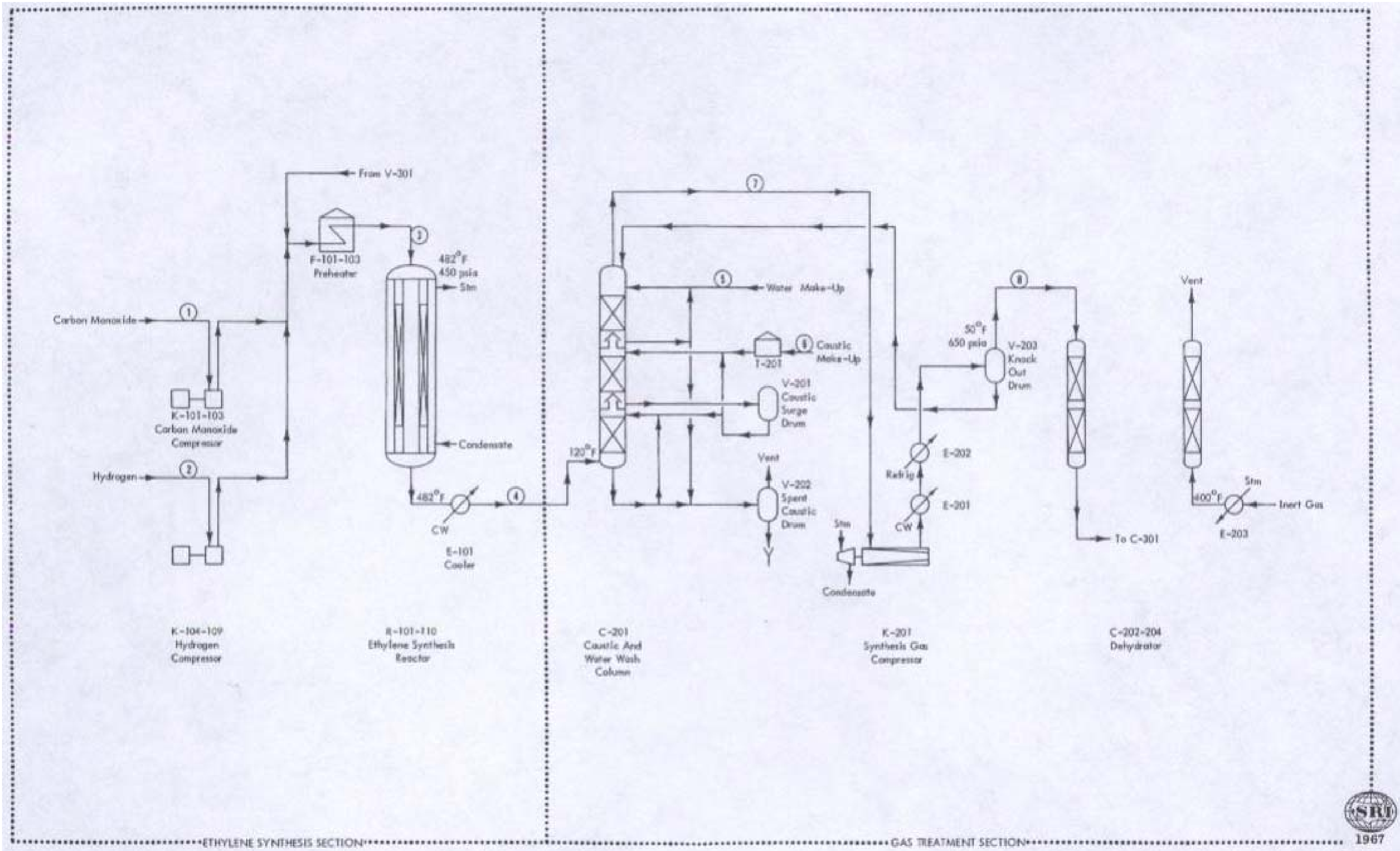

Figure 14: Ethylene synthesis from carbon monoxide and hydrogen (ethylene synthesis and gas treatment sections).

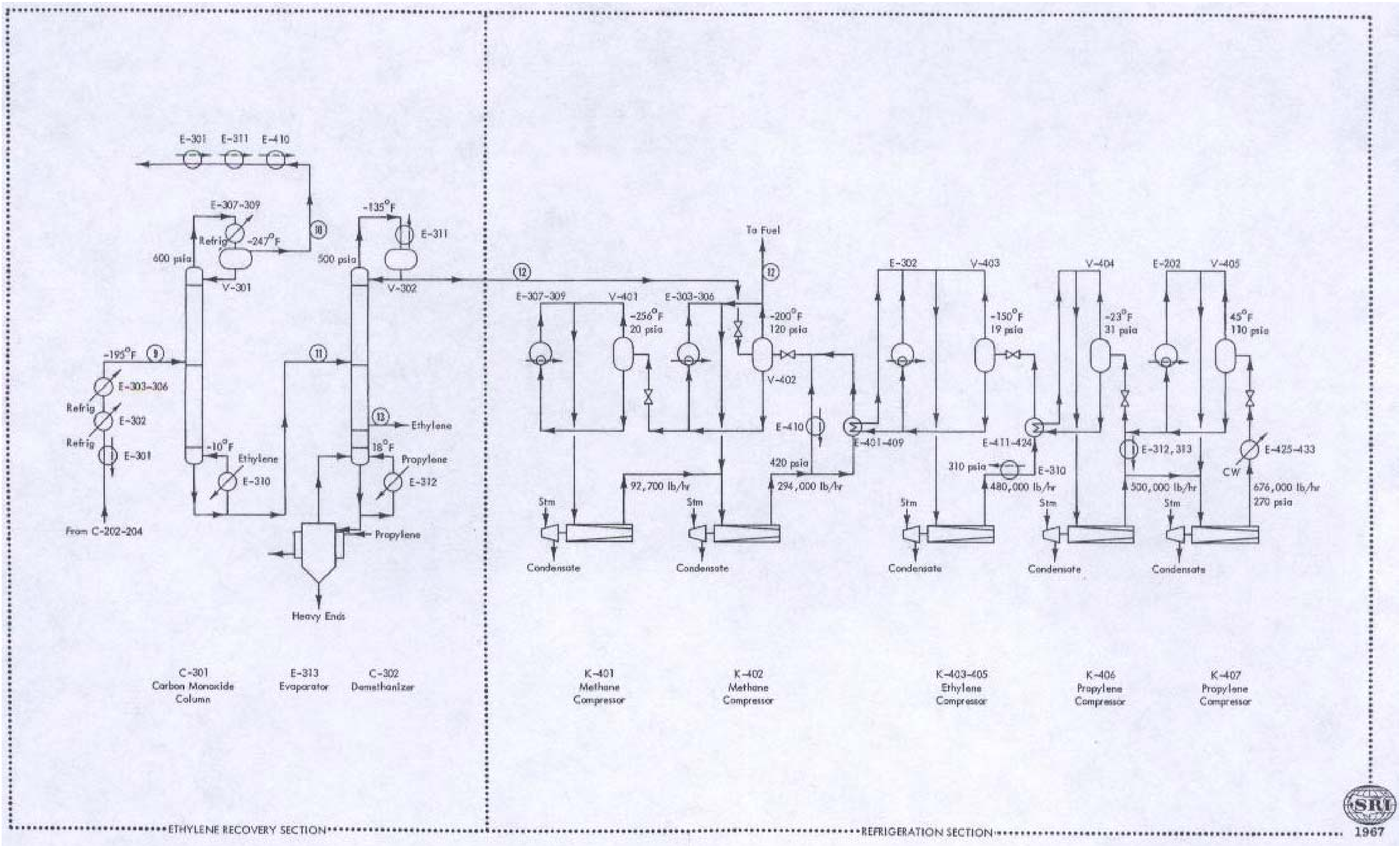

Figure 15: Ethylene synthesis from carbon monoxide and hydrogen (ethylene recovery and refrigeration sections).

Fresh feed of carbon monoxide and hydrogen are compressed separately from atmospheric pressure to 450 psia in a reciprocal compressor. The compressed carbon monoxide and hydrogen gases are mixed with recycled gas, consisting mainly of carbon monoxide and hydrogen and preheated to 482°F. The preheated mixture is fed to the multi-tubular reactor. Heat in the reactor is removed by the generation of steam. Then, the synthesis gas is cooled and charged to the bottom of the caustic and water wash column where carbon dioxide present in the sythesis gas is removed by caustic washing followed by a subsequent water washing. Before passing the synthesis gas to the dehydrator, it is compressed and cooled to condense as much water as possible. Before the dried gas is fed to the carbon monoxide column, it is cooled by three successive stages of heat exchangers. The overhead vapor from the carbon monoxide column is partially condensed by methane refrigerant. All the condensate returns to the column as reflux, and the uncondensed gas is recycled to the ethylene synthesis reactor after three successive heat exchangers. The net bottoms of carbon monoxide are fed to the demethanizer, where methane is removed as overhead product and polymer-grade ethylene is taken as a side stream near the bottom. The overhead vapor condensed by heat exchange with the net overhead stream forms the carbon monoxide column feed, and condensing propylene refrigerant is used as the reboiling medium. Any polymer or heavy ends are concentrated in the bottom of the column and removed using a thin film evaporator.

2.6 Direct Conversion

Natural gas is a source of clean burning fuel that could be used in many petrochemical industries. The conversion of natural gas into hydrocarbon liquids has been a technological goal for many years. Gas to liquid GTL as presented in the previous section depends upon FT process. Unfortunately, this process is complicated and requires many unit operations to achieve its purpose. As a result the plants are rather expensive and require a relatively high price for oil to be competitive.[20] At Texas A&M University a group of researchers have developed an entirely different approach for the conversion of natural gas into hydrocarbon liquids, without producing an intermediate synthetic gas. In the following section, I discuss the new technology of producing ethylene directly from natural gas, known as the GTE process. Today, the opportunity exists for ethylene based upon feed from natural gas.

2.6.1 Natural Gas

As a feedstock, natural gas yields ethylene from its ethane or propane content and forms the basis of a massive chemical industry. Large reserves exist in many regions of the world. Much of the natural gas appears in regions that are remote from markets or pipe lines, and it is called "stranded" gas, which is a natural gas field that has been discovered, but remains unusable for either physical or economic reasons. Most of this gas is flared, re-circulated back into oil reservoirs or not produced. In addition, natural gas has a major disadvantage in transportation. Because of the low density of natural gas, pipeline construction is very expensive; therefore, the drive to develop a new process with improved liquefaction technologies, compression technologies, and gas to liquid technologies. Another technology that would meet with quick adoption is a mean to "activate" the methane in natural gas.[21] Producing ethylene from methane in addition to, or instead of, ethane would greatly increase the available feedstock, because it would avoid the cost of ethane separation from natural gas. Producing hydrogen along with the ethylene could also be a benefit for the chemical industry. While the actual kinetics mechanism is very complex, according to GTE researchers, the overall cracker is:

$$2CH_4 \rightarrow C_2H_6 + H_2 \quad (9)$$

$$C_2H_6 \rightarrow C_2H_4 + H_2 \quad (10)$$

$$C_2H_4 \rightarrow C_2H_2 + H_2 \quad (11)$$

$$C_2H_2 \rightarrow 2C + H_2 \quad (12)$$

Quenching the reaction before the final step in the overall reaction would result in the following reaction:

$$2CH_4 \rightarrow C_2H_2 + 3H_2 \quad (13)$$

The process is essentially two reaction steps and one separation step to produce ethylene. The simplified flow diagram of the process is figure 15.

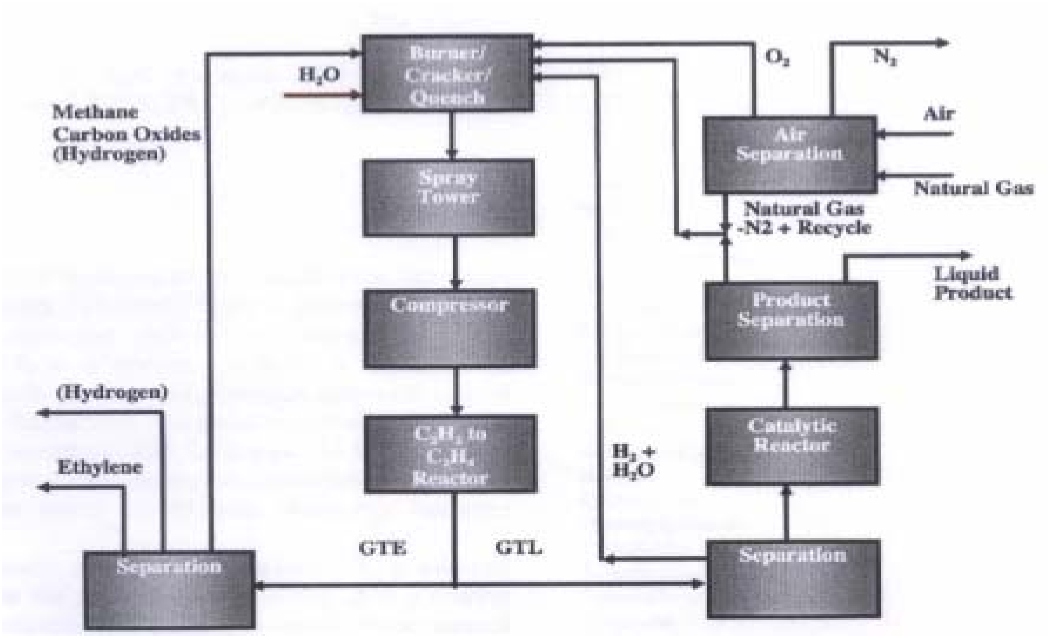

Figure 16: simplified flow diagram of the GTE process (K.R. Hall/ Catalysis Today 106 (2005) 243-246).

The feed gas is treated to remove most of the heavies and sweetened to remove hydrogen sulfide. The process starts by preheating the inlet gas mix with hydrogen recycle stream to about 1000 °F. A portion of the gas fuels an internal combustion thermal cracker into which the reminder of the inlet gas flows. The combustion reaches a temperature sufficient to crack the natural gas hydrocarbon into olefins, primarily acetylene. For a lean gas, the residence time would be relatively longer, the temperature would be relatively higher, and acetylene would be the dominant C_2 product. However, for relatively rich gas, 10% mole C_2+, the residence time would be relatively shorter, and the temperature would be relatively lower.[20] Like other processes of ethylene production, steam is added to the cracker to reduce coke formation, which is possible when high temperature is used to crack the hydrocarbons. The reaction is quenched with water at the exit of the reactor. The cracked gas is fed to the compression section and then it is fed into the hydrogenation section. The flow diagram of the processes is divided into 5 sections: cracking, compression, hydrogenation, amine treatment, and ethylene purification. Figures 17, 18, 19, 20, 21 contain more details for each section.

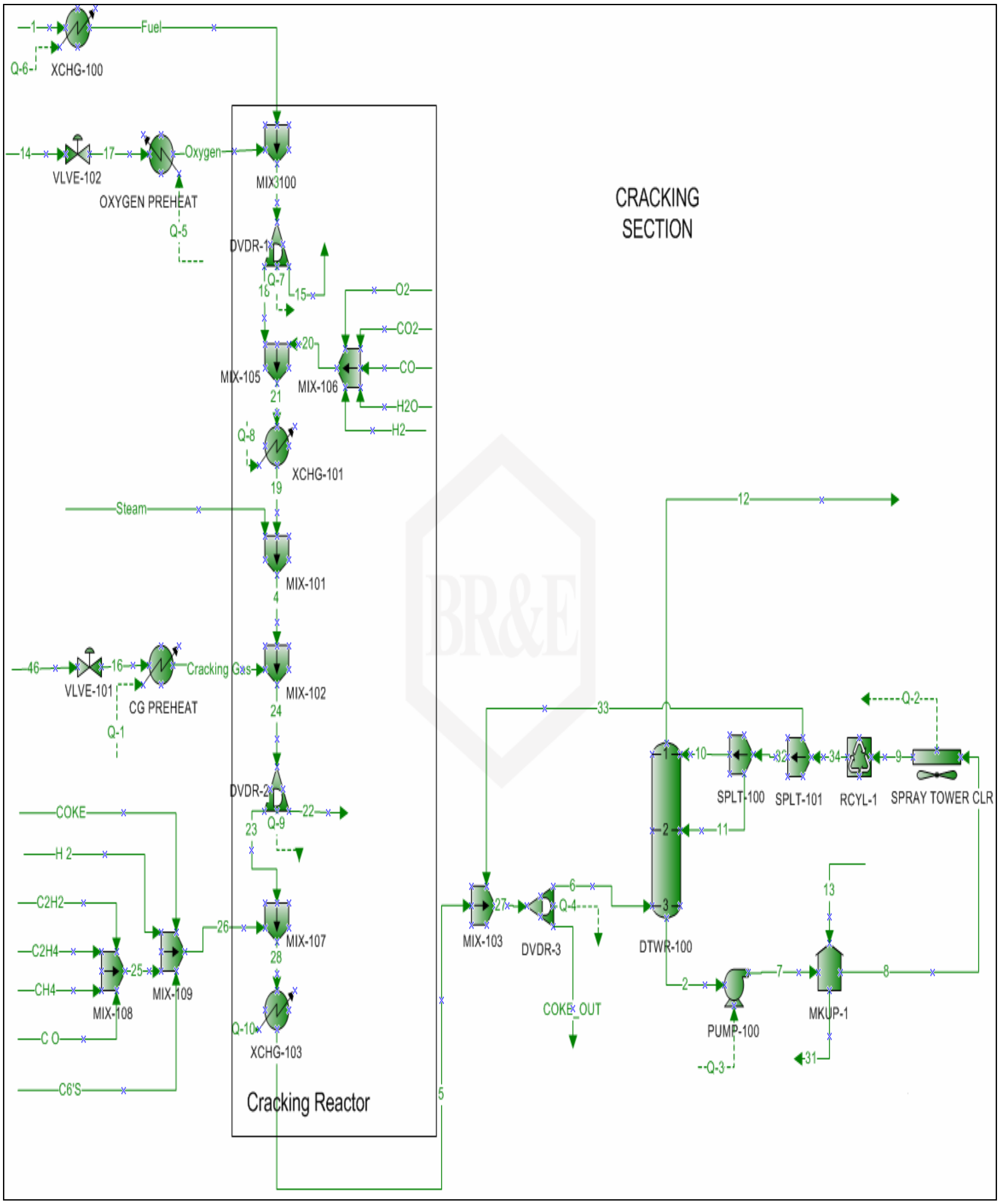

Figure 17: Cracking section of GTE process.

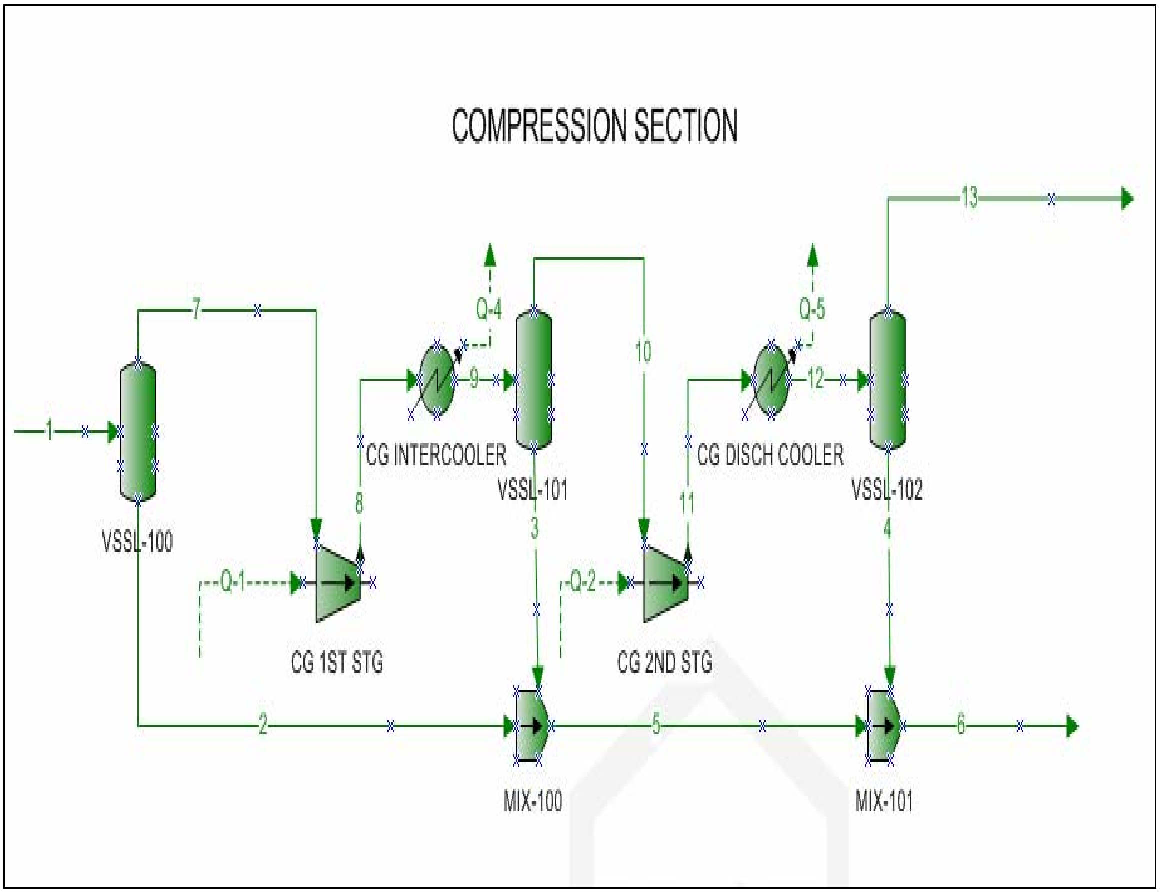

Figure 18: Compression section of GTE process.

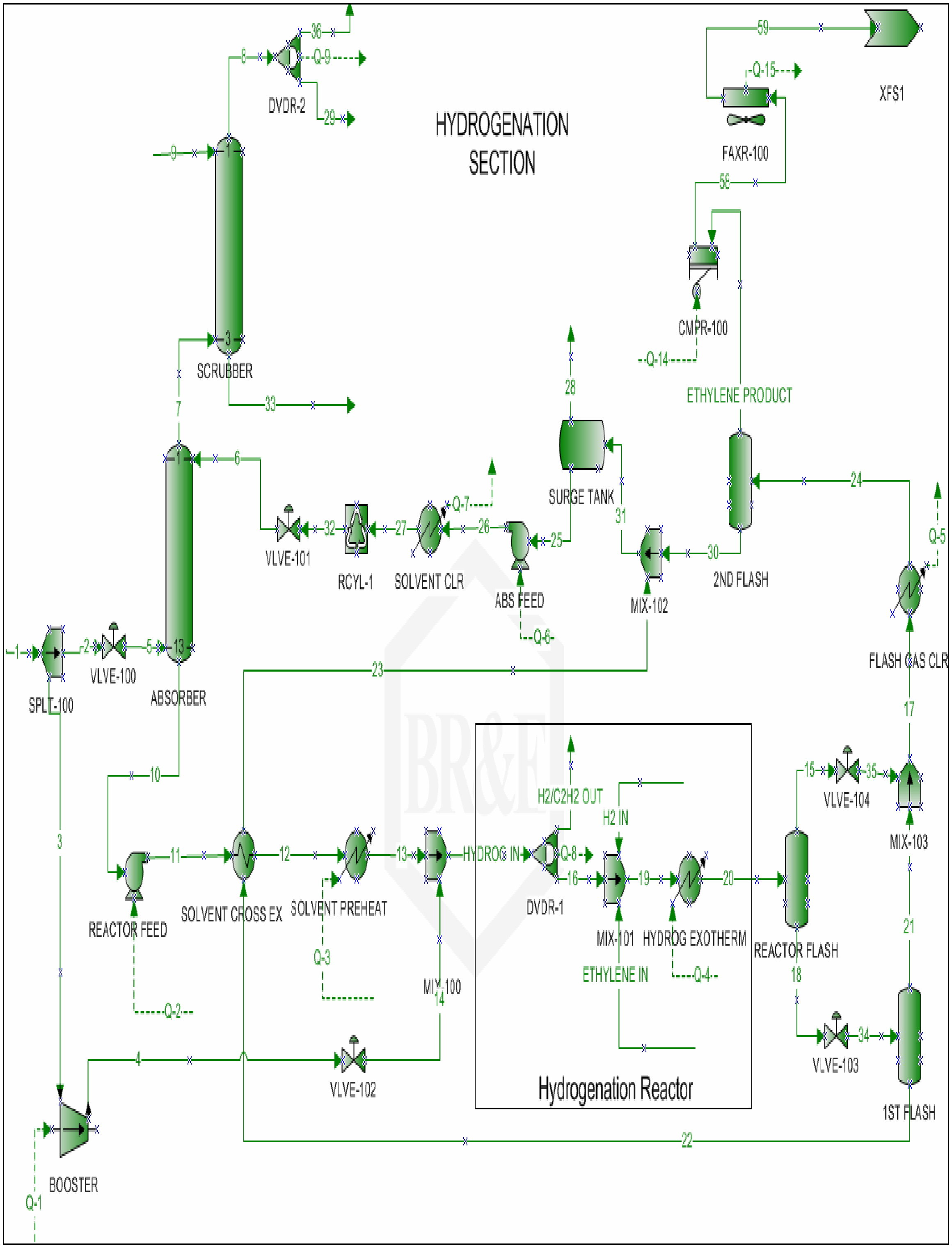

Figure 19: Hydrogenation section of GTE process.

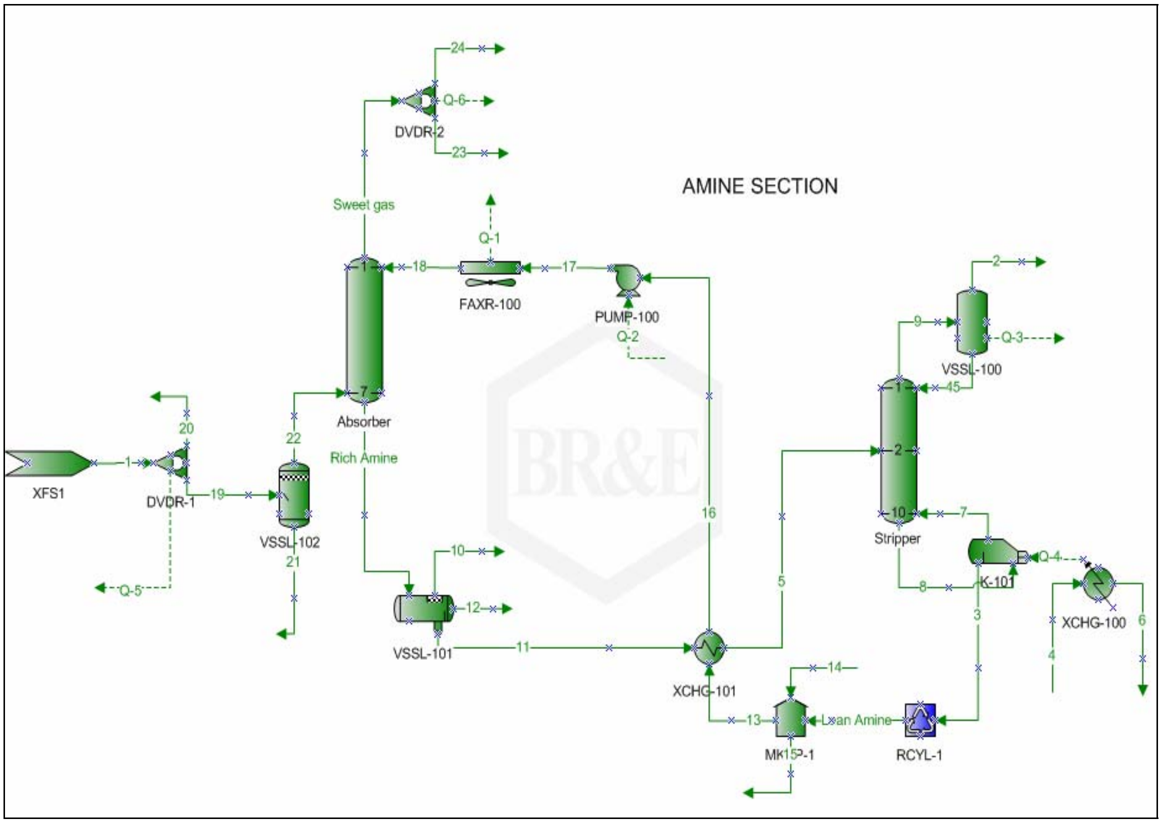

Figure 20: Amine section of GTE process.

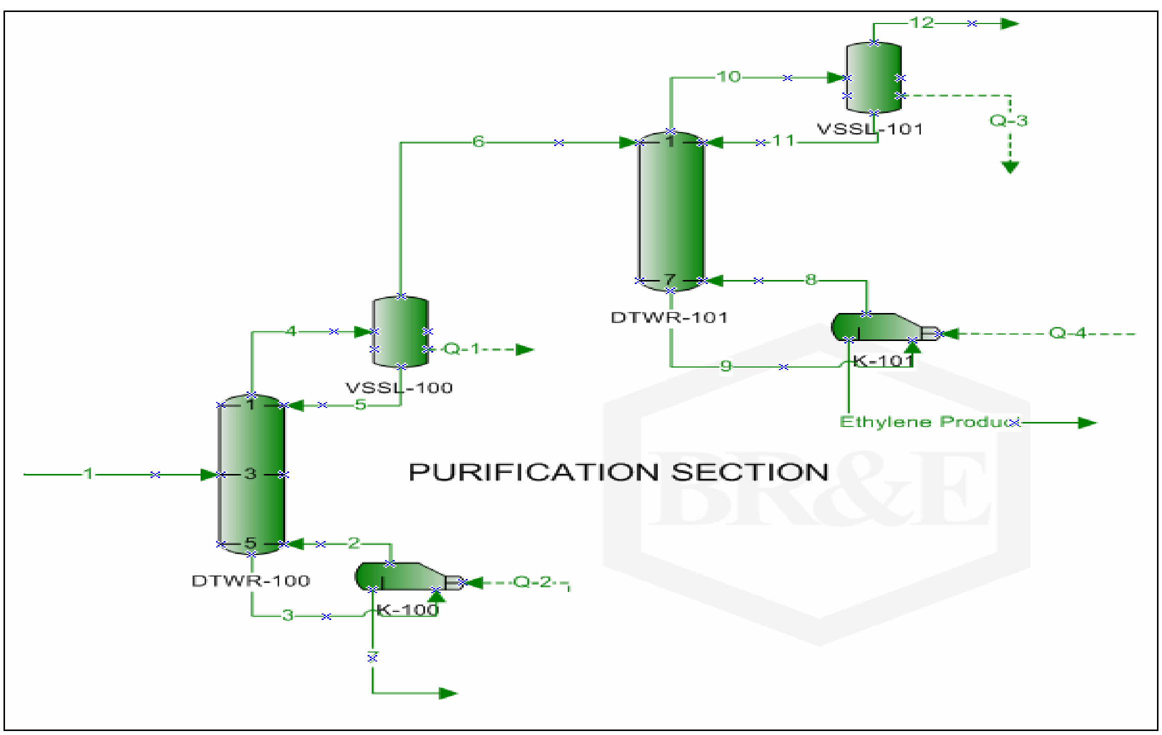

Figure 21: Purification section of GTE process.

Natural gas and recycled fuel are separately fed into the cracker. Steam is also fed to the reactor to decrease coke formation in the cracker. The cracked gas leaves the furnace reactor at 2680 °F and 5 psig. The gas is then quenched with cooling water and enters a molecular sieve dryer. Then, the cracked gas from thermal cracking section enters the compression section, where it is compressed from 15 psig to 138 psig. The compressor section has two stages; each stage consists of vessel, compressor, and heat exchanger. The vessels separate the two phase gas and remove water from the gas, while the compressors increase the pressure of the gas, and the heat exchangers decrease the temperature of the gas.

After removing water between compression stages, the cracked gas enters the hydrogenation section, where the cracked gas is split into two streams. The first stream enters the bottom of the absorber, whereas the second stream enters the hydrogenation reactor. The first stream enters the absorber and then the scrubber to remove acetylene gas. A solvent is used to remove acetylene. After removing acetylene, the overhead vapor of the scrubber is recycled back into the thermal cracker as a fuel. Before entering the hydrogenation reactor, the second stream of the cracked gas is compressed to 280 psig. The hydrogenation reactor removes hydrogen and acetylene, and generates ethylene. The two phase effluent stream exiting the hydrogenation reactor goes into several flash tanks to separate the liquid, which is recycled back to the absorber, and the vapor, which contains ethylene product.

Then, the ethylene product is compressed and cooled before entering ethylene separation and purification. The first unit in ethylene production is an amine unit to remove carbon dioxide. The amine unit consists of two columns: an absorber and a stripper. Ethanolamine solvent the absorber removes carbon dioxide from the ethylene product stream. The sweet gas leaving the absorber is sent to the ethylene purification section, while the bottom rich amine stream contains carbon dioxide that enters the stripper where it is further separated from other components and then flared. Finally, the sweet gas exiting the amine unit enters the purification section, where ethylene is separated and purified to about 99%. The purification section consists of two distillation columns. In the first column, propane and heavier components are separated from ethylene, and they leave in the net bottom stream. The condenser of the first column feeds some of the overhead vapor back to the top of the column, while the reboiler vaporizes some of the bottom product and sends it back the bottom of the column. Similarly, the second column separate ethylene from other components. Liquid ethylene is recovered in the bottom of the column, whereas other components are recovered in the overhead vapor.

3. PROPOSED WORK

Section 2 presents the different processes of ethylene production in commercial use today. Petrochemical companies have many alternatives to produce ethylene. Many factors affect the choice of the process. For example, the cost of the process, the availability of raw material, and environmental issues are some important factors in choosing the process. Economics is the most dominant factor for commercializing any chemical product. The cost of raw materials, utilities, and operations are the main aspects of the economics of chemical processes. My goal is to compare the economics of different ethylene production process including the most recent process that converts natural gas directly to ethylene, GTE. I have evaluated the economics of all process discussed in section 2 and compared the cost of producing ethylene using ethane, propane, naphtha, crude oil, syngas, and natural gas as raw materials. Promax and Capcost software package simulate and evaluate GTE process. Additionally, the cost index provides an update cost estimate for other ethylene production processes. In the following section, I discuss the methodology used to evaluate each process present the results of the economic analysis. The work was broken into two parts.

3.1 Methodology

To achieve the objective of this work, first a literature study was done on the GTE process to develop a better understanding of the new idea behind the process. In addition, I looked at the most common ethylene production process in the literature. Because ethylene production is an old process, I found much information about ethylene processes. Since 1930's, when ethylene was used in chemical industries in United States, many modifications and developments have been introduced to the process in order to increase the production rate and purity of the product and to decrease the utility and operation costs. I focused more upon the general processes and the economics of each method. My main reference was a study by Stanford Research Institute (SRI) about the economics of different ethylene production processes, but I also used other books and articles to understand the ethylene process better. Then, I completed a flow sheet diagram developed by Mr. Joel G. Cantrell, who worked on the gas to ethylene process GTE project. This project proposed a new method for ethylene production using direct conversion of natural gas to liquid hydrocarbons. After completing the flow sheet, I evaluated the economics of GTE process using Capcost. Finally, I changed the cost of the ethylene process reported in SRI in 1967 into the cost of the processes in 2005 using cost index calculator.

3.1.1 Complete GTE Flow Sheet

The researchers at Texas A&M University, Drs. Kenneth Hall, Aydin Akgerman, Rayford G. Anthony, Philip T. Eubank, and Jerry A. Bullin, have developed a new process for conversion of natural gas directly into liquid. This new technology has been licensed by Synfuels International of Dallas for commercialization. Using GTE technology, a pilot plant capable of processing 3 kCMD has been built in Bryan Texas. The new technology can produce ethylene and possibly hydrogen as a products or liquids.

Figure 16 presents a schematic diagram of a possible arrangement for the GTE and GTL processes. Specialized software for gas and chemical processes simulation, Promax simulates the GTE process. Promax is a powerful and adaptable process simulation package that engineers use worldwide to design and optimize gas processing, refining and chemical facilities. Totally integrated with Microsoft visio, excel, and word, Promax is a comprehensive tool that significantly extends the capabilities of its predecessors. A simple user interface, multiple flow sheet capabilities, over 50 thermodynamic packages, 2300 pure components, and friendly examples are a few of the features that made Promax my simulator choice for this project.

The initial flow sheet of the GTE process contained four sections: cracking, compression, hydrogenation, and oligiomerization. This flow sheet could produce ethylene and/or hydrocarbon liquids as intended by researchers at Texas A&M University. Yet, ethylene produced was not purified so that many changes and modification should be employed in order to produce high purity ethylene. Consequently, my first task was to complete the flow sheet to produce pure ethylene as a main product. The ethylene product stream in the original flow sheet contains the composition in table 2

Table 2: Composition of ethylene product stream in original flow sheet of GTE process.

Component	Mole %
Water	2.72
Carbon Dioxide	14.80
Carbon Monoxide	15.04
Hydrogen	12.94
Nitrogen	3.64
Methane	13.95
Ethane	0.07
Ethylene	35.17
Propane	0.02
Propylene	0.02
i-Butane	0.23
i-Pentane	0.46
n-Hexane	0.83
NMP	0.11

My plan was to remove carbon dioxide first and then separate ethylene from the heavier components. To achieve that I added an amine unit to remove carbon dioxide gas from the ethylene product stream. The amine unit appears in figure 12. The amine unit consists of absorber, stripper, heat exchangers, and recycle pump. The inlet stream feeds to the bottom of the absorber, while the solvent, 5.72% MEA, feeds to the top. MEA absorbs most of the carbon dioxide in the ethylene product stream and leaves the bottom of the absorber via the amine rich stream, whereas the sweet gas leaves the top of absorber to the separation units. Before going to the stripper, the rich amine stream enters a flash tank where the vapor phase is separated from the liquid phase. The vapor phase contains carbon monoxide and hydrocarbons, while the liquid phase contains carbon dioxide, water, and traces of hydrocarbons. In the stripper the carbon dioxide is separated from other components and sent to vent. Other components exiting the bottom of the stripper are recycled and send back to the absorber through a recycle pump. Before venting carbon dioxide from the top of the stripper column, part of it is condensed and refluxed to the top of the stripper.

The next section that I have added to the flow sheet of GTE process is a separation unit, shown in figure 13. The sweet gas exiting the top of the absorber enters the separation unit. The separation unit consists of two distillation columns. The first distillation separates propane and heavier components from the ethylene product stream. To achieve that, I used a specification option in the column targeting 95 mol% of propane in the bottom of the column. Then, the top stream exiting the first distillation column enters the top of the second distillation column where ethylene product is recovered in the bottom product. The target is to recover pure ethylene from ethane and other gases. To reach my target, I have specified 99 mol% of ethylene in the bottoms. The process appears successful and I have achieved about 99% ethylene in my product stream. I have tried different scenarios to achieve the best possible results by changing the location of the feed stream and manipulating the reflux ratio in the distillation columns. The best result that I have achieved in the final product stream appears in table 3. The results of the entire process and Promax simulations are in appendix A.

Table 3: The composition of the product stream.

Component	Mole %
Carbon Dioxide	0.005
Methane	0.002
Ethane	0.197
Ethylene	99.780
Propane	0.002
Propylene	0.009
i-Butane	0.005

3.1.2 Estimate the cost of the Flow Sheet

After successful simulation of the GTE flow sheet to produce high purity ethylene, the next task is to estimate the cost of the GTE process. A computer program, Capcost, estimates the cost of GTE process. Capcost is a program written in the Microsoft Windows programming environment. The program requires the user to input information about the equipment, for example, the capacity, operating pressure, and materials of construction. Other information such as output file name and number of units, also are required. Before using Capcost, some information is required to estimate the cost of the units. For example, heat transfer area of the heat exchangers, the volume of the flash tanks and vessels, and volume of the reactors are necessary to use Capcost. This information is not provided in the initial GTE flow sheet; therefore, I have sized the heat exchangers and vessels in a separate flow sheet. The cooling media used in heat exchangers is cooling water at 85 °F and atmospheric pressure. The heating media is low and medium pressure steam. The heat transfer area of the heat exchangers is calculated using a rating option in Promax. The input data for rating option is shell inside diameter, number of shell in series or parallel, tube length, number of cross passes, and material of construction. Carbon steel is the material of construction for most of the heat exchangers and other parameters vary depending upon the desired output temperature, pressure, and flow rate. To estimate the heat transfer area in reboilers, the following equation is used:

$$A = \frac{Q}{U \Delta T_{LMTD}} \tag{14}$$

where

Q = power input or heat lost.

U = overall heat transfer coefficient.

ΔT_{LMTD}= log mean temperature difference.

The heat transfer coefficient is estimated based upon the property of the fluid. Table 4 shows the calculated heat transfer area of the reboilers used in the process.

Table 4: Calculated area of heat transfer in the reboilers.

Section	purification	purification	amine
Unit	K-100	K-101	K-101
overall heat transfer coefficient U (Btu/ft^2*hr*F)	125	125	125
heat transferred Q (Btu/hr)	9066078	70657598	89053140
ΔT_{LMTD}	73	182	18
Area of heat transfer (ft^2)	990	3105	40414

Similarly, vessel volume is determined using a sizing option in Promax. The required inputs of the sizing option are type of the vessel, corrosion allowance, shell material of construction, and head material of construction. Corrosion allowance of 0.25 inch is used in all vessels, and carbon steel is the material of construction of almost all vessels. Table 5 shows the calculated volume and dimensions of the vessels.

Table 5: Vessels volume and dimensions in GTE process.

Section	Unit	vessel inside diameter ft	vessel height ft	vessel internal volume ft^3
compression	vssl 100	7	7.4	285
compression	vssl 101	14	13.6	2098
compression	vssl 102	10.5	10.7	927
hydrogenation	reactor flash	19	18.4	5208
hydrogenation	1st flash	19	18.1	5133
hydrogenation	2nd flash	10	10.7	840
Amine	vssl 102	3.5	3.7	36
Amine	veel 101	13	12.8	1703
Amine	vssl 100	4	5.6	70
purification	vssl 100	9	8.3	525
purification	vssl 101	13	14.3	1894

Then, I determine the height and diameter of the column to estimate the cost of the columns in the process. Fluid mass density of the liquid and vapor are provided by Promax, and the following equations determine the diameter and height of the column.

$$u_c = Kv\sqrt{\frac{\rho_L - \rho_V}{\rho_V}} \tag{15}$$

u_c = maximum permissible velocity

K_V = empirical coefficient

ρ_L = liquid density

ρ_V = vapor density

Based upon the ratio of mass flow rate of liquid to vapor and plate spacing, which I have chosen to be 18 inches, the empirical coefficient comes from figure 22.

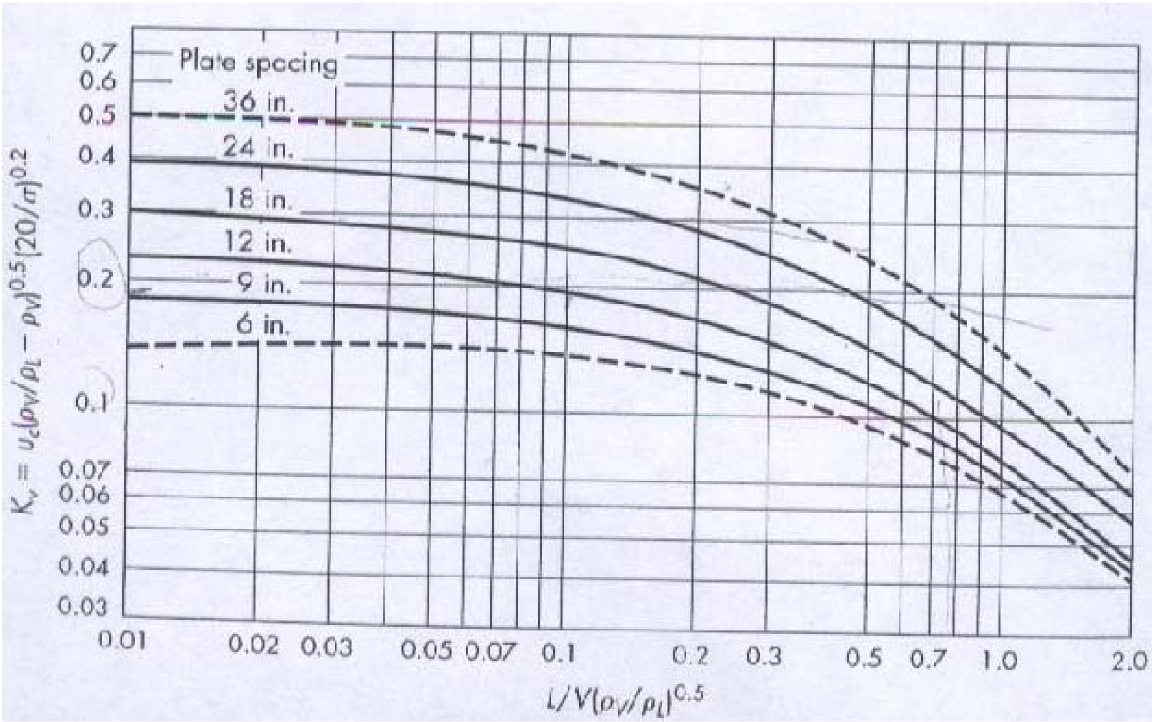

Figure 22: Value of Kv at flooding conditions for sieve plates L/V= ratio mass flow rate liquid to vapor, u is in feet per second, and σ is in dynes per centimeter. [J.R. Fair, Petrol. Chem. Eng., 33(10):45,1961. Courtesy Petroleum Engineering.]

After that, the maximum permissible velocity is calculated. Then, the cross sectional area is calculated by dividing the mass flow rate by the maximum permissibly velocity. Knowing the cross sectional area of the column, I could determine the diameter of the column. The height of the column is based upon the number of stages and spacing between plates. Table 6 shows the calculated diameter and height of each column.

Table 6: Diameter and height of the columns in GTE process.

Section	Unit	diameter (ft)	height (ft)
cracking	DTWR-100	7	13
hydrogenation	absorber	6	28
hydrogenation	scrubber	6	13
Amine	absorber	3	19
Amine	stripper	4	24
Purification	DTWR-100	5	16
purification	DTWR-101	8	19

The reactor is one of the important units in chemical processes, but Promax does not contain this option. In the GTE process, the reactors results from combining separators, dividers, and heat exchangers blocks. An 8 million standard cubic feet per day (MMSCFD) cracking gas reactor has an estimated \$100,000 purchase cost. The cracking gas reactor used in the project has a capacity of 50 MMSCFD. Using the following relationship between the purchased cost and an attribute of the equipment related to units of capacity, the cost of the cracking reactor is estimated.

$$\frac{Ca}{Cb} = \left(\frac{Aa}{Ab}\right)^{0.6} \tag{16}$$

where A = Equipment cost attribute

C = Purchase cost

subscript: a refers to equipment with the required attribute

b refers to equipment with the base attribute

In addition, the cost of hydrogenation reactor is estimated based upon the catalyst used. The density of the catalyst is about 45 lb/ft^3 and the solvent volume to catalyst volume ratio is 10. Tables 7, 8, 9, 10, and 11 present the equipment cost for each section in the GTE process as calculated by Capcost.

Table 7: Equipment costs of cracking section in GTE process.

Unit	Number	Type	Purchase Cost ($)	Bare Module Cost ($)
Heat Exchanger	E-101	Fixed, Sheet, or U-Tube	505,000	1,660,000
Fired Heater	H-101	Pyrolysis Furnace	992,000	2,110,000
Fired Heater	H-102	Pyrolysis Furnace	743,000	1,580,000
Fired Heater	H-103	Pyrolysis Furnace	1,450,000	3,080,000
Pump	P-101	Centrifugal	74,700	297,000
Tower	T-101	3 Carbon Steel Sieve Trays	33,100	110,000
Reactor	Z-101	Cracking Reactor	300,281	1,660,555
Total				10,497,555

Table 8: Equipment costs of compression section in GTE process.

Unit	Number	Type	Purchase Cost ($)	Bare Module Cost ($)
Compressor	C-101	Centrifugal	1,600,000	4,390,000
Compressor	C-102	Centrifugal	5,200,000	14,300,000
Heat Exchanger	E-102	Fixed, Sheet, or U-Tube	125,000	412,000
Heat Exchanger	E-103	Fixed, Sheet, or U-Tube	117,000	385,000
Vessel	V-101	Vertical	199,000	861,000
Vessel	V-102	Vertical	138,000	639,000
Vessel	V-103	Vertical	48,400	432,000
			Total	21,419,000

Table 9: Equipment costs of hydrogenation section in GTE process.

Unit	Number	Type	Purchase Cost ($)	Bare Module Cost ($)
Compressor	C-103	Centrifugal	642,000	1,760,000
Compressor	C-104	Centrifugal	3,010,000	8,250,000
Heat Exchanger	E-104	Fixed, Sheet, or U-Tube	435,000	1,450,000
Heat Exchanger	E-105	Fixed, Sheet, or U-Tube	120,000	393,000
Heat Exchanger	E-106	Fixed, Sheet, or U-Tube	30,400	101,000
Heat Exchanger	E-107	Fixed, Sheet, or U-Tube	105,000	348,000
Pump	P-102	Centrifugal	75,500	347,000
Pump	P-103	Centrifugal	103,000	447,000
Tower	T-102	13 Carbon Steel Sieve Trays	43,100	148,000
Tower	T-103	3 Carbon Steel Sieve Trays	17,800	79,000
Vessel	V-104	Vertical	162,000	2,050,000
Vessel	V-105	Vertical	235,000	958,000
Vessel	V-106	Vertical	43,100	175,000
Reactor	Z-102	Hydrogenation Reactor	166,256	919,397
			Total	4,776,397

Table 10: Equipment costs of amine section in GTE process.

Unit	Number	Type	Purchase Cost ($)	Bare Module Cost ($)
Heat Exchanger	E-108	Fixed, Sheet, or U-Tube	34,900	115,000
Heat Exchanger	E-109	Fixed, Sheet, or U-Tube	136,000	453,000
Heat Exchanger	E-110	Kettle Reboiler	3,680,000	12,100,000
Pump	P-104	Centrifugal	44,500	211,000
Tower	T-104	7 Carbon Steel Sieve Trays	12,500	48,900
Tower	T-105	10 Carbon Steel Sieve Trays	21,300	89,100
Vessel	V-107	Horizontal	2,840	11,700
Vessel	V-108	Horizontal	52,100	582,000
Vessel	V-109	Horizontal	4,470	13,400
			Total	13,056,100

Table 11: Equipment costs of purification section in GTE process.

Unit	Number	Type	Purchase Cost ($)	Bare Module Cost ($)
Heat Exchanger	E-111	Kettle Reboiler	89,800	307,000
Heat Exchanger	E-112	Kettle Reboiler	282,000	964,000
Tower	T-106	5 Carbon Steel Sieve Trays	16,500	91,700
Tower	T-107	7 Carbon Steel Sieve Trays	42,700	292,000
Vessel	V-110	Vertical	18,400	171,000
Vessel	V-111	Vertical	66,700	769,000
			Total	2,594,700

From Figure 23, we see that the most expensive section of GTE process is the compression section, and the cheapest section is the purification section. This result from the fact that compressors are very expensive and the volume of the cracked gas entering the compression section, which compresses the gas from 15psig to 135 psig, is very large (about 3.36 million ft^3/hr). The purification section is the cheapest section because I have considered the last two distillation columns of the process in the purification section. Furthermore, it was relatively easy separation because the fed gas has had most of the impuritie removed in the previous steps.

In addition to the equipment cost, Capcast program also calculated the utility of the GTE process based on the heating and cooling media of each unit. Table 12 shows the utility cost of the process. Figure 24 shows the annual cost distribution of utility cost in the five sections of the process. Cracking section uses up 71% of the total annual utility cost of the GTE plant, because the natural gas cracks at very high temperature, which requires large heat duty in the furnaces.

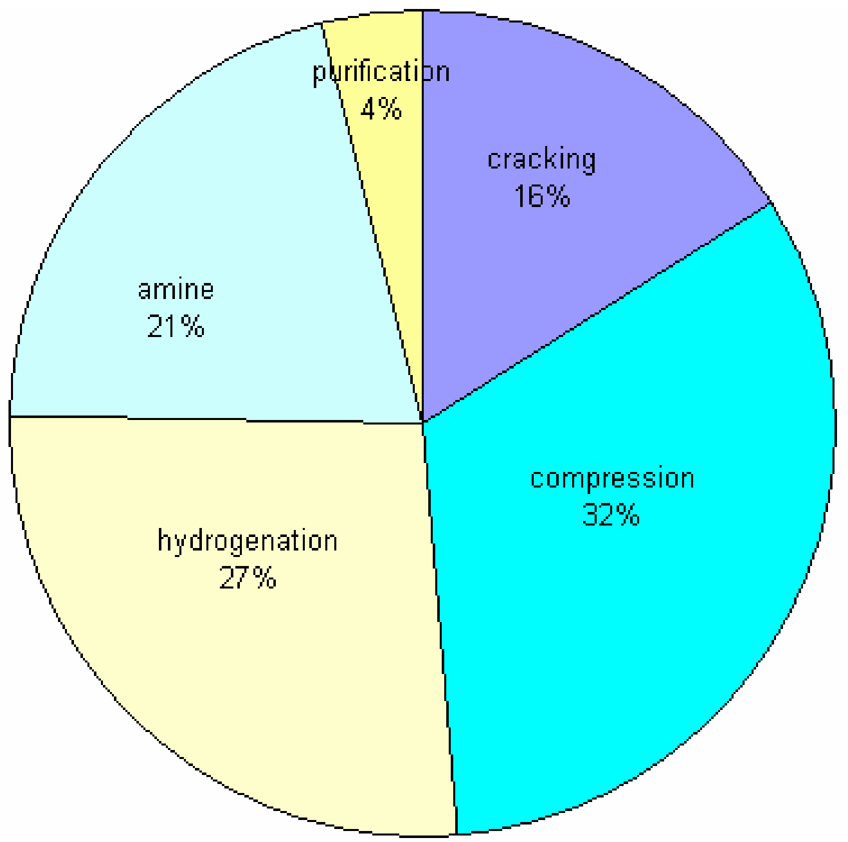

Figure 23: Equipment cost distribution in GTE process.

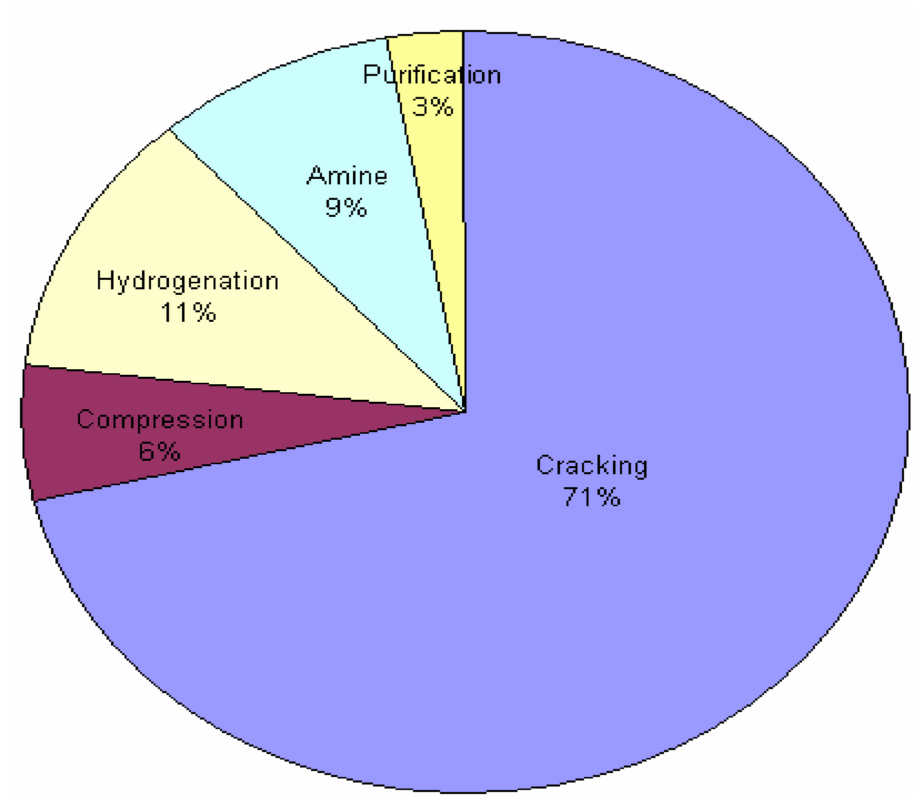

Figure 24: Annual utility cost of GTE plant.

Table 12: Utility cost of the GTE process.

Section	Unit	Total Module Cost	Annual Utility Cost
Cracking	E-101	1,962,122	209,400,000
Compression	E-102	490,000	8,400,000
Compression	E-103	455,000	9,200,000
Hydrogenation	E-104	1,705,000	20,900,000
Hydrogenation	E-105	460,000	2,510,000
Hydrogenation	E-106	120,000	4,600,000
Hydrogenation	E-107	411,000	6,200,000
Amine	E-108	135,000	10,800,000
Amine	E-109	534,000	11,300,000
Amine	E-110	14,274,000	4,800,000
Purification	E-111	362,000	550,000
Purification	E-112	1,137,000	8,100,000
Fired Heater	H-101	2,490,000	2,320,000
Fired Heater	H-102	1,870,000	1,620,000
Fired Heater	H-103	3,630,000	3,700,000
Cracking	P-101	351,000	226,000
Hydrogenation	P-102	409,000	229,000
Hydrogenation	P-103	527,000	343,000
Purification	P-104	249,000	143,000
Total Annual Cost			305,341,000

Other costs such as service facilities, total fixed capital start up, and working capital, labor, and all other costs were estimated using the same method that SRI report used.

3.1.3 Change the Cost of Other Methods

Following the estimation of GTE equipment and utility costs, the next step is to compare the cost of GTE process with the costs other processes of ethylene production. To estimate the cost of other processes, I have used the economic report on ethylene from SRI on 1967. This report evaluated the most common methods of ethylene production and provides some information about the processes. To compare the economics of these processes to GTE process, first I have converted these costs to 2005 value. It is essential to update these costs to take into account changing economic conditions such as inflation. This update uses.[22]

$$C_2 = C_1\left(\frac{I_2}{I_1}\right) \tag{17}$$

where: C = purchase Cost

I = cost Index

subscripts: 1 refers to base time when cost is known

2 refers to time when cost is desired

After updating the cost of ethylene plants, I matched the plant capacities of these plants with the GTE plant using the plant capacity exponent provided in SRI report for each method. The plant capacity of other processes is 500 million pounds of ethylene per year, whereas the plant capacity of the GTE process I have simulated is 321 million pounds of ethylene per year. I have used the following expression to convert plant costs of other processes such as fixed capital cost, and process unit costs based upon 500 million pounds of ethylene per year into 321 million pounds of ethylene per year

$$C_{2005} = \left(\frac{A_{2005}}{A_{1967}}\right)(C_{1967})^n \tag{18}$$

Where C_{2005} = cost at 2005

C_{1967} = cost at 1967

A_{1967} = plant capacity of 500 pounds of ethylene per year

A_{2005} = plant capacity of 321 pounds of ethylene per year

n = plant capacity exponent

Tables 13 and 14 summarize the total capital investment of ethylene production processes using different feedstocks based upon a plant capacity of 321 million ethylene pounds per year in 2005. All costs are reported per pound of ethylene. In GTE process economics, I have used the same methodology and standards as those in the SRI reports for labor, direct and indirect costs of the plant. Old costs and more detailed costs are in appendix B.

Table 13: Summary of total fixed capital of different ethylene production processes.

Feedstock	Process Unit Investment ($)	Utilities & Tankage ($)	Total Process & Utilities cost ($)	General Service Facilities at 15% of Above ($)	Total Fixed Capital ($)
Ethane	62,078,637	10,869,398	72,948,035	10,942,205	83,890,240
Propane	68,653,462	12,733,101	81,386,563	12,207,984	93,594,548
Naphtha	71,218,425	13,975,418	85,193,843	12,779,076	97,972,920
Crude Oil	79,605,322	23,130,285	102,735,607	15,410,341	118,145,948
CO+H2	86,083,590	9,565,759	95,649,349	14,347,402	109,996,752
NG	65,560,752	13,112,150	78,672,902	11,800,935	90,473,838

Table 14: Summary of total fixed investment of different ethylene production processes.

Feedstock	Interest on Construction Loan at 7%/yr ($)	Start-up Costs ($)	Working Capital ($)	Total Capital Investment, Not Including Land ($)
Ethane	5,872,317	8,390,410	9,070,831	107,223,798
Propane	6,551,618	9,358,893	10,955,383	120,460,442
Naphtha	6,858,104	9,798,125	11,922,966	126,552,115
Crude Oil	8,270,216	11,814,171	13,034,000	151,264,335
CO+H2	7,699,773	11,000,413	7,444,602	136,141,539
NG	6,333,169	9,047,384	9,952,122	115,806,512

3.2 Economical Analysis

The economic analysis of ethylene production processes using different feedstocks suggests that the most economic feedstock is crude oil. However, using natural gas as a feedstock via GTE is more economical than other processes that are used today to produce ethylene. Table 15 is a summary of the economic analyses of the most common ethylene presses. All costs are based upon pounds of ethylene produced.

Table 15: Summary of ethylene product cost by different feedstocks.

Feedstock	Ethane	Propane	Naphtha	Crude Oil	CO + H2	Natural Gas
Total Labor ¢	0.679	0.740	0.764	0.843	0.903	0.712
Total Materials ¢	7.385	12.970	23.184	21.151	36.196	2.711
Total Utilities¢	3.224	3.870	4.277	6.036	4.938	2.714
Total Direct Operating Costs ¢	11.288	17.581	28.226	28.030	42.037	6.137
Total Indirect Operating Costs ¢	3.840	4.286	4.486	5.322	4.962	4.128
Total By-Product Credit ¢	1.646	10.833	18.288	25.389	0.312	0
Net Production Cost ¢	13.48	11.03	14.42	7.96	46.69	10.26

The table demonstrates that the labor and indirect cost of all the processes are very close. However, the utilities, raw material, and operating costs of the GTE process are the lowest among all processes. This result is logical, because natural gas price today is much cheaper than the feedstocks used in other processes. For example, the cost of ethane and propane should be more than the cost of natural gas because they come from natural gas. Further, the costs of naphtha and crude oil are expensive because the price of oil has increasing dramatically in the last few years. Synthesis gas is the most expensive process because of the costly process of producing carbon monoxide and hydrogen. Crude oil is the most economical process because of the by products of this processes beside ethylene. For example, the byproducts of crude oil are propylene, light oil, butylene-butane, ethane, propane, residue gas, naphthalene, and others; whereas, the byproduct of synthesis gas is methane only and I assume no byproduct for GTE. The future feedstock of ethylene as well as many other chemical processes seems to be natural gas because of the abundant amount of natural gas that has not been utilized in the past. The amount of gas reserves is estimated to be about 1,190.62 trillion cubic feet as of January 2000.[23] In addition, natural gas is a clean-burning gas that has a less negative affect on the environment. Further, the price of oil is increasing and does not give the impression that the era of cheap oil may return. With the increase of oil demand and oil reservoir conditions, scientists predict the end of oil is likely to be in decades, therefore, they are looking for other sources of energy. Countries rich in natural gas, such as Russia, United States, Iran, and Qatar are investing in natural gas processes today to use their natural gas reserves better.

4. CONCLUSIONS

The objective of this work is to evaluate the economics of a new process for ethylene production proposed by a group of researchers at Texas A&M University. The process converts natural gas directly to ethylene without making syn gas as an intermediate step. The gas to ethylene process (GTE) consists of two reaction sections and one separation section to produce ethylene from natural gas. With high temperature and careful attention to residence time natural gas can be cracked to ethylene. Some catalysts are also employed in the GTE process to produce ethylene or liquid hydrocarbons from natural gas. The literature contains many processes used to produce ethylene. The most common feedstocks for ethylene production are ethane, propane, naphtha, crude oil, and synthesis gas (carbon monoxide and hydrogen). The economics of all these processes have been reviewed and compared to the GTE economics. I also have examined the general processes of ethylene production using these different feedstocks.

Promax and Capcost software were used to simulate and evaluate the economics of the GTE process. The cost index and some expressions were also employed to update the cost and adjust the plant capacity of the different ethylene processes to match the GTE process. In term of capital investment costs, the compression section was the most expensive in the GTE processes. However, for utility costs, the cracking section was most expensive.

The results show that ethylene production by natural gas is more economical than the other processes, except crude oil. The reasons for natural gas superiority are low cost raw material and simplicity of the process. In addition, natural gas is more abundant and environmentally safer than other processes; therefore, it seems to be a better energy and feedstock source than crude oil in the future. Still, more studies and optimization of the current process could further lower the cost of utilities and operation of the GTE process.

LITERATURE CITED

1. Shigeyoshi Takaoka. “Ethylene” Stanford Research Institute. *Process Economics Program, report No. 29*, 1967: August.

2. Hall K.R. A new gas to liquids (GTL) or gas to ethylene (GTE) technology, *Catalysis Today*. **106**:243-246(2005).

3. Kniel L., Winter O., Stork K. Ethylene: *Keystone to the Petrochemical Industry*, Marcel Dekker, Inc., New York, (1978).

4. Chang T. Ethylene-Capacity Growth Slowed in 1998. *Oil & Gas Journal*. **97** (13), 50, (1999).

5. Waller D. Ethylene from NGL Feedstocks-Part 2: World ethylene outlook. *Hydrocarbon Process*. **62**: 11, 187 (1983).

6. The Innovation Group. Chemical Profiles Available on the web at http://www.the-innovation-group.com/ChemProfiles/Ethylene.htm. Accessed February 18, 2007.

7. Kolmetz.com. Technical Articles Available on the web at http://kolmetz.com/Article-039.htm, Accessed February 18, 2007.

8. AbdulRahman A. AlRabiah. “Design and Optimization of Ethylene Production Utilizing New Technologies”. PhD. Dissertation, University of Colorado, Boulder, (2001).

9. Picciotti M. Ethylene technologies-I: Novel ethylene route are available, but economics inhibit their use. *Oil & Gas Journal*. **95**: 25,53 (1997).

10. Picciotti M. Ethylene technologies-conclusion: New technologies developing, but steam cracking remains king. *Oil & Gas Journal*. **95**: 26, 71 (1997).

11. Chemical Market Association, Inc. Chemical prices Available on the web at http://www.cmaiglobal.com/. Accessed February 18, 2007.

12. Lam W. K., Mehra, Y. R., Mullins D. Advanced Ethylene Process. Presented at *AICHE Spring National Meeting*. Houton:1993.

13. Hoeermann H. *Chem. Ind*. **21**, 860 (1969).

14. Mosberg E.*Chem.-Ing.-Tech*. **43**: 3,131 (1971).

15. Davis H.G., Keister R.G. The Advaced Cracking Reactor (ACR): A Process for Cracking Hydrocarbon Liquids at Short residence Times, High Temperatures, and Law Partial Pressure, Presented at *ACS Meeting*, Philadelphia, 1975; April 6-11.

16. Germain .E. Thermal decomposition of pure methane at 1000° in a pipe still. *Genie Chem*. **88**: 122-7 (1962).

17. Rice F. O. Thermal decomposition of organic compounds from standpoint of free radicals. I. saturated hydrocarbons. *J. Am. Chem. Soc*. **53**: May, 1959-72. (1931).

18. Sherwood, P. W. Production of ethylene from petroleum sources, Part I. *Petrol. Refiner.* **30**: September: 220-5. (1951).

19. Sherwood, P. W. New advances in ethylene production, part I. *Petroleum*. **19**: May:161-4. (1956).

20. Martin R. Thermal decomposition of propane. I. Introduction and experimental study of the pyrolysis of pure propane. *J. Chem. Phys*. **3**: 286-97. (1964).

21. Wikipedia.org . Available on the web at http://en.wikipedia.org. Accessed at February 23, 2007.

22. Hall K.R. A radically new process to convert natural gas into hydrocarbon liquids, in *AIChE National Meeting*. Houston: 2001.

23. Peters Max S., Timmerhaus Klaus D. *Plant Design and Economics for Chemical Engineers*. Fourth Edition, McGraw Hill, New York (1991).

24. NaturalGas.org. Chemical Prices Available on the web at http://www.naturalgas.org/overview/resources.asp. Accessed at February 7, 2007.

SUPLEMMENTARY MATERAILS CONSULTED

Hall K.R. New low-cost GTL naphtha scheme debuts. *GTL News.* (2000) :August.

Hall K.R. A new route from natural gas to liquid fuels. *Chem. Eng.* (2001) :June.

Hall K.R. New GTL method for fewer heavy fractions. *Hydrocarbon Process.* (2002): November.

McCabe Warren L., Smith Julian C., Harriott Peter, *Unit Operations of Chemical Engineering*. Sixth Edition, McGrew Hill, New York, (2001).

Turton Richard, Balilie Richard C., Whiting Wallace B., Shaeiwitz, Joseph A. *Analysis, Synthesis, and Design of Chemical Processes*. Second Edition, Pearson Hall PTR, Boston (2003).

APPENDIX A

Table A.1: Summary of different ethylene production processes based on 500 million lb/yr of ethylene on 1967.

plant capacity : 500 million lb/yr of ethylene	fired tubular heater			autothermic cracking in fluidized bed	synthesis
feed stock	ethane	propane	naphtha	crude oil	CO+H2
capital costs ($)					
process unit	14,324,000	15,911,000	16,725,000	18,860,000	20,941,000
utilities and tankage	2,508,000	2,951,000	3,282,000	5,480,000	2,327,000
fixed capital	19,357,000	21,691,000	23,008,000	27,991,000	26,758,000
total capital cost, not including land	24,744,000	27,918,000	29,719,000	35,838,000	33,112,000
plant capacity exponent	0.68	0.69	0.72	0.74	0.8
production cost (¢/lb)					
labor	0.151	0.161	0.165	0.178	0.191
raw and process materials					
ethane @0.9¢/lb	1.12				
propane @0.9¢/lb		2.064			
naphtha @1.35¢/lb			3.804		
minas crude @$2.2/bbl (.74¢/lb)				3.301	
oxygen @0.9¢/lb				1.634	
CO @1.9¢/lb					4.052
H2 @6.3¢/lb					1.961
others	0.043	0.045	0.047	0.153	0.041
subtotal	1.163	2.109	3.851	5.088	6.054
total materials	1.255	2.211	3.957	5.207	6.185
utilities	0.565	0.678	0.748	1.063	0.863
total direct operating costs	1.971	3.05	4.87	6.448	7.239
indirect costs	0.819	0.937	1.011	1.152	1.166
total production costs	2.79	3.987	5.881	7.6	8.405
by product credit					
proplene @2.8¢/lb		0.805	1.39	1.515	
pyrolysis gasoline @1.34¢/lb	0.055	0.359	0.552	0.827	
C4 fraction @2.5¢/lb		0.162	0.737	0.689	
naphthalene fraction @3.5¢/lb				0.625	
residue gas @ 34¢/million btu	0.175	0.522	0.4	0.391	0.053
others	0.052		0.046	0.308	
subtotal	0.282	1.848	3.125	4.355	0.053
net production cost	2.508	2.139	2.756	3.245	8.352
confidence rating	B	B	B	C	D

Table A.2: Summary of different ethylene production processes based on 321 million lb/yr of ethylene on 1967.

plant capacity : 321,868,872 million lb/yr of ethylene	fired tubular heater			autothermic cracking in fluidized bed	synthesis
feed stock	ethane	propane	naphtha	crude oil	CO+H2
capital costs ($)					
process unit	10,616,623	11,741,043	12,179,700	13,614,018	14,721,925
utilities and tankage	1,858,873	2,177,601	2,390,061	3,955,717	1,635,926
fixed capital	14,346,969	16,006,219	16,755,189	20,205,195	18,811,387
total capital cost, not including land	18,339,691	20,601,246	21,642,362	25,869,522	23,278,371
plant capacity exponent	0.68	0.69	0.72	0.74	0.8
production cost (¢/lb)					
labor	0.151	0.161	0.165	0.178	0.191
raw and process materials					
ethane @0.9¢/lb	1.12				
propane @0.9¢/lb		2.064			
naphtha @1.35¢/lb			3.804		
minas crude @$2.2/bbl (.74¢/lb)				3.301	
oxygen @0.9¢/lb				1.634	
CO @1.9¢/lb					4.052
H2 @6.3¢/lb					1.961
others	0.043	0.045	0.047	0.153	0.041
subtotal	1.163	2.109	3.851	5.088	6.054
total materials	1.255	2.211	3.957	5.207	6.185
utilities	0.565	0.678	0.748	1.063	0.863
total direct operating costs	1.971	3.05	4.87	6.448	7.239
indirect costs	0.819	0.937	1.011	1.152	1.166
total production costs	2.79	3.987	5.881	7.6	8.405
by product credit					
proplene @2.8¢/lb		0.805	1.39	1.515	
pyrolysis gasoline @1.34¢/lb	0.055	0.359	0.552	0.827	
C4 fraction @2.5¢/lb		0.162	0.737	0.689	
naphthalene fraction @3.5¢/lb				0.625	
residue gas @ 34¢/million btu	0.175	0.522	0.4	0.391	0.053
others	0.052		0.046	0.308	
subtotal	0.282	1.848	3.125	4.355	0.053
net production cost	2.508	2.139	2.756	3.245	8.352
confidence rating	B	B	B	C	D

Table A.3: Summary of different ethylene production processes based on 321 million lb/yr of ethylene on 2005.

plant capacity : 321,868,872 million lb/yr of ethylene	fired tubular heater			autothermic cracking in fluidized bed	synthesis
feed stock	ethane	propane	naphtha	crude oil	CO+H2
capital costs ($)					
process unit	62,078,637	68,653,462	71,218,425	79,605,322	86,083,590
utilities and tankage	10,869,398	12,733,101	13,975,418	23,130,285	9,565,759
fixed capital	83,891,109	93,593,253	97,972,707	118,145,948	109,995,925
total capital cost, not including land	107,237,774	120,461,777	126,549,500	151,266,995	136,115,744
plant capacity exponent	0.68	0.69	0.72	0.74	0.8
production cost (¢/lb)					
labor	0.679	0.740	0.764	0.843	0.903
raw and process materials					
ethane @5.26¢/lb	6.549				
propane @5.26¢/lb		12.061			
naphtha @7.89¢/lb			22.234		
minas crude @$12.86/bbl (4.33¢/lb)				19.316	
oxygen @5.26¢/lb				0.094	
CO @11.11¢/lb					23.687
H2 @36.84¢/lb					11.457
others	0.250	0.260	0.270	0.890	0.240
subtotal	6.799	12.321	22.504	20.301	35.384
total materials	7.385	12.970	23.140	30.450	36.170
utilities	3.224	3.870	4.277	6.036	4.938
total direct operating costs	11.288	17.581	28.226	28.030	42.037
indirect costs	3.840	4.286	4.486	5.322	4.962
total production costs	15.128	21.867	32.712	33.352	46.999
by product credit					
proplene @16.37¢/lb		4.731	8.136	8.856	
pyrolysis gasoline @7.84¢/lb	0.321	2.101	3.230	4.837	
C4 fraction @14.62¢/lb		0.950	4.313	3.962	
naphthalene fraction @20.4¢/lb				3.644	
residue gas @ 7.08¢/million btu	1.020	3.051	2.339	2.291	0.312
others	0.300		0.270	1.800	
subtotal	1.641	10.833	18.288	25.391	0.312
net production cost	13.487	11.034	14.424	7.961	46.687
confidence rating	B	B	B	C	D

Table A.4: Capital costs of different ethylene production processes based on 500 million lb/yr of ethylene on 1967.

	Ethane	Propane	Naphtha	Crude Oil	CO+H2
process unit investment	$14,324,000	$15,911,000	$16,725,000	$18,860,000	$20,941,000
Utilities & tankage	$2,508,000	$2,951,000	$3,282,000	$5,480,000	$2,327,000
Total process & utilities cost	$16,832,000	$18,862,000	$20,007,000	$24,340,000	$23,268,000
General service facilities at 15% of above	$2,524,800	$2,829,300	$3,001,050	$3,651,000	$3,490,200
Total fixed capital	$19,356,800	$21,691,300	$23,008,050	$27,991,000	$26,758,200
interest on construction loan at 7%/yr	$1,354,976	$1,518,391	$1,610,564	$1,959,370	$1,873,074
start-up costs	$1,936,000	$2,169,000	$2,301,000	$2,799,000	$2,676,000
working capital	$2,093,000	$2,539,000	$2,800,000	$3,088,000	$1,811,000
total capital investment, not including land	$24,740,776	$27,917,691	$29,719,614	$35,837,370	$33,118,274

Table A.5: Capital costs of different ethylene production processes based on 321 million lb/yr of ethylene on 1967.

	Ethane	Propane	Naphtha	Crude Oil	CO+H2
process unit investment	$10,616,623	$11,741,043	$12,179,700	$13,614,018	$14,721,925
Utilities & tankage	$1,858,873	$2,177,601	$2,390,061	$3,955,717	$1,635,926
Total process & utilities cost	$12,475,496	$13,918,644	$14,569,761	$17,569,735	$16,357,850
General service facilities at 15% of above	$1,871,324	$2,087,797	$2,185,464	$2,635,460	$2,453,678
Total fixed capital	$14,346,821	$16,006,441	$16,755,225	$20,205,195	$18,811,528
interest on construction loan at 7%/yr	$1,004,277	$1,120,451	$1,172,866	$1,414,364	$1,316,807
start-up costs	$1,434,919	$1,600,548	$1,675,665	$2,020,447	$1,881,279
working capital	$1,551,284	$1,873,578	$2,039,053	$2,229,061	$1,273,168
total capital investment, not including land	$18,337,301	$20,601,018	$21,642,809	$25,869,067	$23,282,782

Table A.6: Capital costs of ethylene production based on ethane, propane, and naphtha feedstock on 2005 and plant capacity of 321 million lb/yr of ethylene.

	Ethane	Propane	Naphtha
process unit investment	$62,078,637	$68,653,462	$71,218,425
Utilities & tankage	$10,869,398	$12,733,101	$13,975,418
Total process & utilities cost	$72,948,035	$81,386,563	$85,193,843
General service facilities at 15% of above	$10,942,205	$12,207,984	$12,779,076
Total fixed capital	$83,890,240	$93,594,548	$97,972,920
interest on construction loan at 7%/yr	$5,872,317	$6,551,618	$6,858,104
start-up costs	$8,390,410	$9,358,893	$9,798,125
working capital	$9,070,831	$10,955,383	$11,922,966
total capital investment, not including land	$107,223,798	$120,460,442	$126,552,115

Table A.7: Capital costs of ethylene production based on crude oil, synthesized gas, and natural gas feedstock on 2005 and plant capacity of 321 million lb/yr of ethylene.

	Crude Oil	CO+H2	NG
process unit investment	$79,605,322	$86,083,590	$65,560,752
Utilities & tankage	$23,130,285	$9,565,759	$13,112,150
Total process & utilities cost	$102,735,607	$95,649,349	$78,672,902
General servicce facilities at 15% of above	$15,410,341	$14,347,402	$11,800,935
Total fixed capital	$118,145,948	$109,996,752	$90,473,838
interest on construction loan at 7%/yr	$8,270,216	$7,699,773	$6,333,169
start-up costs	$11,814,171	$11,000,413	$9,047,384
working capital	$13,034,000	$7,444,602	$9,952,122
total capital investment, not including land	$151,264,335	$136,141,539	$115,806,512

Table A.8: Operation costs of ethylene production based on ethane feedstock on 1967.

labor			
operating	4.4 $/hr	7 operators/shift	0.054
maintenance	0.03 of total unit investment		0.086
control laboratory	0.2 of operating labor		0.011
total labor (¢)			0.151
materials			
raw & process		untis/lb	¢/lb
Ethane	0.9 ¢/lb	1.245 lb	1.121
caustic soda	2.9 ¢/lb	0.006 lb	0.017
hydrogentation catalyst	1.5 $/lb	0.000144 lb	0.022
desiccant	1.36 $/lb	0.0000266 lb	0.004
maintenance	3%yr of process unit investment		0.086
operating	10% of operating labor		0.005
Total materials			1.254
untilities		untis/lb	¢/lb
CW	0.002 ¢/gal	42.006 gal	0.084
Steam	0.055 ¢/lb	4.317 lb	0.237
Electricity	1 ¢/kwh	0.017 kw	0.017
Process water	0.03 ¢/gal	0.047 gal	0.001
Fuel	0.000034 ¢/btu	6607 btu	0.225
Total Utilities			0.564
Total Direct Operating costs			1.970
plant overhead	0.8 % of total labor	0.121	
taxes & insurance	0.02 % of fixed capital	0.077	
plant cost		2.168	
sales & research	0.06 % of sales	NA	
cash expenditures		2.168	
depreciation	0.1 %/yr of fixed capital	0.387	
interest on working capital	0.06 %/yr	0.025	
total product cost		2.580	
Indirect costs			0.610
by-product credit		untis/lb	¢/lb
residue gas	1.21 ¢/lb	0.144 lb	0.174
pyrolsis gasoline	1.34 ¢/lb	0.041 lb	0.055
C3-C4 fraction	0.9 ¢/lb	0.058 lb	0.052
			0.281
Net Production Cost			2.299

Table A.9: Operation costs of ethylene production based on propane feedstock on 1967.

labor					
operating	4.4	$/hr	7	operators/shift	0.054
maintenance	0.03	of total unit investment			0.095
control laboratory	0.2	of operating labor			0.011
total labor					0.160
materials					
raw & process			untis/lb		¢/lb
propane	0.9	¢/lb	2.293	lb	2.064
caustic soda	2.9	¢/lb	0.006	lb	0.017
hydrogentation cat	1.5	$/lb	0.000168	lb	0.025
desiccant	1.36	$/lb	0.0000148	lb	0.002
maintenance	3%yr of process unit investment				0.095
operating	10% of operating labor				0.005
Total materials					2.209
untilities			untis/lb		¢/lb
CW	0.002	¢/gal	50.312	gal	0.101
Steam	0.055	¢/lb	4.861	lb	0.267
Electricity	1	¢/kwh	0.018	kw	0.018
Process water	0.03	¢/gal	0.066	gal	0.002
Fuel	0.000034	¢/btu	8514	btu	0.289
Total Utilities					0.677
Total Direct Operating costs					3.047
plant overhead	0.8	% of total labor	0.128		
taxes & insurance	0.02	% of fixed capital	0.087		
plant cost			3.262		
sales & research	0.06	% of sales	NA		
cash expenditures			3.262		
depreciation	0.1	%/yr of fixed capital	0.434		
interest on working capital	0.06	%/yr	0.030		
total product cost			3.726		
Indirect costs					0.679
by-product credit			untis/lb		¢/lb
propylene	2.8	¢/lb	0.289	lb	0.809
pyrolsis gasoline	1.34	¢/lb	0.268	lb	0.359
butylene-butanes	2.5	¢/lb	0.065	lb	0.163
residue gas	0.78	¢/lb	0.669	lb	0.522
					1.853
Net Production Cost					1.873

Table A.10: Operation costs of ethylene production based on naphtha feedstock on 1967.

labor					
operating	4.4	$/hr	7	operators/shift	0.054
maintenance	0.03	of total unit investment			0.100
control laboratory	0.2	of operating labor			0.011
total labor					0.165
materials					
raw & process			untis/lb		¢/lb
light naphtha	1.35	¢/lb	2.818	lb	3.804
caustic soda	2.9	¢/lb	0.008	lb	0.023
hydrogentation cat	1.5	$/lb	0.000156	lb	0.023
desiccant	1.36	$/lb	0.000007	lb	0.001
maintenance	3%yr of process unit investment				0.100
operating	10% of operating labor				0.005
Total materials					3.958
untilities			untis/lb		¢/lb
CW	0.002	¢/gal	57.02	gal	0.114
Steam	0.055	¢/lb	3.199	lb	0.176
Electricity	1	¢/kwh	0.02	kw	0.020
Process water	0.03	¢/gal	0.066	gal	0.002
Fuel	0.000034	¢/btu	12835	btu	0.436
Total Utilities					0.748
Total Direct Operating costs					4.871
plant overhead	0.8	% of total labor	0.132		
taxes & insurance	0.02	% of fixed capital	0.092		
plant cost			5.095		
sales & research	0.06	% of sales	NA		
cash expenditures			5.095		
depreciation	0.1	%/yr of fixed capital	0.460		
interest on working capital	0.06	%/yr	0.034		
total product cost			5.589		
Indirect costs					0.718
by-product credit			untis/lb		¢/lb
propylene	2.8	¢/lb	0.497	lb	1.392
pyrolsis gasoline	1.34	¢/lb	0.412	lb	0.552
butylene-butanes	2.5	¢/lb	0.295	lb	0.738
fuel oil	0.66	¢/lb	0.07	lb	0.046
residue gas	0.78	¢/lb	0.513	lb	0.400
					3.128
Net Production Cost					2.461

Table A.11: Operation costs of ethylene production based on crude oil feedstock on 1967.

labor					
operating	4.4	$/hr	7	operators/shift	0.054
maintenance	0.03	of total unit investment			0.113
control laboratory	0.2	of operating labor			0.011
total labor					0.178
raw & process			untis/lb		¢/lb
Crude Oil	0.74	¢/lb	4.461	lb	3.301
oxygen (98%)	0.9	¢/lb	1.816		1.634
Diethanolamine	20	¢/lb	0.005		0.100
caustic soda	2.9	¢/lb	0.01	lb	0.029
hydrogentation cat	1.5	$/lb	0.000254	lb	0.038
desiccant	1.36	$/lb	0.0000162	lb	0.002
maintenance	3%yr of process unit investment				0.113
operating	10% of operating labor				0.005
Total materials					5.223
untilities			untis/lb		¢/lb
CW	0.002	¢/gal	105.5	gal	0.211
Steam	0.055	¢/lb	12.7	lb	0.699
Electricity	1	¢/kwh	0.038	kw	0.038
Process water	0.03	¢/gal	0.123	gal	0.004
Fuel	0.000034	¢/btu	3382	btu	0.115
Total Utilities					1.066
Total Direct Operating costs					6.467
plant overhead	0.8	% of total labor	0.142		
taxes & insurance	0.02	% of fixed capital	0.112		
plant cost			6.722		
sales & research	0.06	% of sales	NA		
cash expenditures			6.722		
depreciation	0.1	%/yr of fixed capital	0.560		
interest on working capital	0.06	%/yr	0.037		
total product cost			7.319		
Indirect costs					0.851
by-product credit			untis/lb		¢/lb
propylene	2.8	¢/lb	0.541	lb	1.515
light oil	1.34	¢/lb	0.617	lb	0.827
butylene-butanes	2.5	¢/lb	0.271	lb	0.678
ethane	0.69	¢/lb	0.227	lb	0.157
propane	0.68	¢/lb	0.059	lb	0.040
residue gas	0.36	¢/lb	1.086	lb	0.391
coke	0.47	¢/lb	0.236	lb	0.111
naphthalene fracti	3.5	¢/lb	0.178	lb	0.623
					4.341
Net Production Cost					2.978

Table A.12: Operation costs of ethylene production based on carbon monoxide and hydrogen feedstock on 1967.

labor					
operating	4.4	$/hr	7	operators/shift	0.054
maintenance	0.03	of total unit investment			0.126
control laboratory	0.2	of operating labor			0.011
total labor					0.190
materials					
raw & process			untis/lb		¢/lb
CO	1.9	¢/lb	2.132	lb	4.051
H2	6.3	¢/lb	0.311	lb	1.959
caustic soda	2.9	¢/lb	0.01	lb	0.029
hydrogentation cat	1.5	$/lb	0.0000056	lb	0.001
desiccant	1.36	$/lb	0.000084	lb	0.011
maintenance	3%yr of process unit investment				0.126
operating	10% of operating labor				0.005
Total materials					6.182
untilities			untis/lb		¢/lb
CW	0.002	¢/gal	60.322	gal	0.121
Steam	0.055	¢/lb	5.689	lb	0.313
Electricity	1	¢/kwh	0.355	kw	0.355
Process water	0.03	¢/gal	0.114	gal	0.003
Fuel	0.000034	¢/btu	2097	btu	0.071
Total Utilities					0.863
Total Direct Operating costs					7.236
plant overhead	0.8	% of total labor	0.152		
taxes & insurance	0.02	% of fixed capital	0.107		
plant cost			7.495		
sales & research	0.06	% of sales	NA		
cash expenditures			7.495		
depreciation	0.1	%/yr of fixed capital	0.535		
interest on working capital	0.06	%/yr	0.022		
total product cost			8.052		
Indirect costs					0.816
by-product credit			untis/lb		¢/lb
methane	0.73	¢/lb	0.073	lb	0.053
					0.053
Net Production Cost					7.999

Table A.13: Operation costs of ethylene production based on ethane feedstock on 2005.

labor					
operating	4.4	$/hr	7	operators/shift	0.084
maintenance	0.03	of total unit investment			0.579
control laboratory	0.2	of operating labor			0.017
total labor					0.679
materials					
raw & process			untis/lb		¢/lb
Ethane	5.26	¢/lb	1.245	lb	6.549
caustic soda	16.96	¢/lb	0.006	lb	0.102
hydrogentation cat	8.77	$/lb	0.000144	lb	0.126
desiccant	7.95	$/lb	0.0000266	lb	0.021
maintenance	3%yr of process unit investment				0.579
operating	10% of operating labor				0.008
Total materials					7.385
untilities			untis/lb		¢/lb
CW	0.01	¢/gal	42.006	gal	0.420
Steam	0.32	¢/lb	4.317	lb	1.381
Electricity	5.85	¢/kwh	0.017	kw	0.099
Process water	0.18	¢/gal	0.047	gal	0.008
Fuel	0.000199	¢/btu	6607	btu	1.315
Total Utilities					3.224
Total Direct Operating costs					11.288
plant overhead	0.8	% of total labor	0.543		
taxes & insurance	0.02	% of fixed capital	0.521		
plant cost			12.353		
sales & research	0.06	% of sales	NA		
cash expenditures			12.353		
depreciation	0.1	%/yr of fixed capital	2.606		
interest on working capital	0.06	%/yr	0.169		
total product cost			15.128		
Indirect costs					3.840
by-product credit			untis/lb		¢/lb
residue gas	7.08	¢/lb	0.144	lb	1.020
pyrolsis gasoline	7.84	¢/lb	0.041	lb	0.321
C3-C4 fraction	5.26	¢/lb	0.058	lb	0.305
					1.646
Net Production Cost					13.482

Table A.14: Operation costs of ethylene production based on propane feedstock on 2005.

labor					
operating	4.4	$/hr	7	operators/shift	0.084
maintenance	0.03	of total unit investment			0.640
control laboratory	0.2	of operating labor			0.017
total labor					0.740
materials					
raw & process			untis/lb		¢/lb
propane	5.26	¢/lb	2.293	lb	12.061
caustic soda	16.96	¢/lb	0.006	lb	0.102
hydrogentation cat	8.77	$/lb	0.000168	lb	0.147
desiccant	7.95	$/lb	0.0000148	lb	0.012
maintenance	3%yr of process unit investment				0.640
operating	10% of operating labor				0.008
Total materials					12.970
untilities			untis/lb		¢/lb
CW	0.01	¢/gal	50.312	gal	0.503
Steam	0.32	¢/lb	4.861	lb	1.556
Electricity	5.85	¢/kwh	0.018	kw	0.105
Process water	0.18	¢/gal	0.066	gal	0.012
Fuel	0.000199	¢/btu	8514	btu	1.694
Total Utilities					3.870
Total Direct Operating costs					17.581
plant overhead	0.8	% of total labor	0.592		
taxes & insurance	0.02	% of fixed capital	0.582		
plant cost			18.755		
sales & research	0.06	% of sales	NA		
cash expenditures			18.755		
depreciation	0.1	%/yr of fixed capital	2.908		
interest on working capital	0.06	%/yr	0.204		
total product cost			21.867		
Indirect costs					4.286
by-product credit			untis/lb		¢/lb
propylene	16.37	¢/lb	0.289	lb	4.731
pyrolsis gasoline	7.84	¢/lb	0.268	lb	2.101
butylene-butanes	14.62	¢/lb	0.065	lb	0.950
residue gas	4.56	¢/lb	0.669	lb	3.051
					10.833
Net Production Cost					11.034

Table A.15: Operation costs of ethylene production based on naphtha feedstock on 2005.

labor			
operating	4.4 $/hr	7 operators/shift	0.084
maintenance	0.03 of total unit investment		0.664
control laboratory	0.2 of operating labor		0.017
total labor			0.764
materials			
raw & process		untis/lb	¢/lb
light naphtha	7.89 ¢/lb	2.818 lb	22.234
caustic soda	16.96 ¢/lb	0.008 lb	0.136
hydrogentation cat	8.77 $/lb	0.000156 lb	0.137
desiccant	7.95 $/lb	0.000007 lb	0.006
maintenance	3%yr of process unit investment		0.664
operating	10% of operating labor		0.008
Total materials			23.184
untilities		untis/lb	¢/lb
CW	0.01 ¢/gal	57.02 gal	0.570
Steam	0.32 ¢/lb	3.199 lb	1.024
Electricity	5.85 ¢/kwh	0.02 kw	0.117
Process water	0.18 ¢/gal	0.066 gal	0.012
Fuel	0.000199 ¢/btu	12835 btu	2.554
Total Utilities			4.277
Total Direct Operating costs			28.226
plant overhead	0.8 % of total labor	0.612	
taxes & insurance	0.02 % of fixed capital	0.609	
plant cost		29.446	
sales & research	0.06 % of sales	NA	
cash expenditures		29.446	
depreciation	0.1 %/yr of fixed capital	3.044	
interest on working capital	0.06 %/yr	0.222	
total product cost		32.712	
Indirect costs			4.486
by-product credit		untis/lb	¢/lb
propylene	16.37 ¢/lb	0.497 lb	8.136
pyrolsis gasoline	7.84 ¢/lb	0.412 lb	3.230
butylene-butanes	14.62 ¢/lb	0.295 lb	4.313
fuel oil	3.86 ¢/lb	0.07 lb	0.270
residue gas	4.56 ¢/lb	0.513 lb	2.339
			18.288
Net Production Cost			14.424

Table A.16: Operation costs of ethylene production based on crude oil feedstock on 2005.

labor					
operating	4.4	$/hr	7	operators/shi ft	0.084
maintenance	0.03	of total unit investment			0.742
control laboratory	0.2	of operating labor			0.017
total labor					0.843
raw & process			untis/lb		¢/lb
Crude Oil	4.33	¢/lb	4.461	lb	19.316
oxygen (98%)	0.052	¢/lb	1.816		0.094
Diethanolamine	116.95	¢/lb	0.005		0.585
caustic soda	16.96	¢/lb	0.01	lb	0.170
hydrogentation cat	8.77	$/lb	0.000254	lb	0.223
desiccant	7.95	$/lb	0.0000162	lb	0.013
maintenance	3%yr of process unit investment				0.742
operating	10% of operating labor				0.008
Total materials					21.151
untilities			untis/lb		¢/lb
CW	0.01	¢/gal	105.5	gal	1.055
Steam	0.32	¢/lb	12.7	lb	4.064
Electricity	5.85	¢/kwh	0.038	kw	0.222
Process water	0.18	¢/gal	0.123	gal	0.022
Fuel	0.000199	¢/btu	3382	btu	0.673
Total Utilities					6.036
Total Direct Operating costs					28.030
plant overhead	0.8	% of total labor	0.674		
taxes & insurance	0.02	% of fixed capital	0.734		
plant cost			29.438		
sales & research	0.06	% of sales	NA		
cash expenditures			29.438		
depreciation	0.1	%/yr of fixed capital	3.671		
interest on working capital	0.06	%/yr	0.243		
total product cost			33.352		
Indirect costs					5.322
by-product credit			untis/lb		¢/lb
propylene	16.37	¢/lb	0.541	lb	8.856
light oil	7.84	¢/lb	0.617	lb	4.837
butylene-butanes	14.62	¢/lb	0.271	lb	3.962
ethane	4.03	¢/lb	0.227	lb	0.915
propane	3.98	¢/lb	0.059	lb	0.235
residue gas	2.11	¢/lb	1.086	lb	2.291
coke	2.75	¢/lb	0.236	lb	0.649
naphthalene fracti	20.47	¢/lb	0.178	lb	3.644
					25.389
Net Production Cost					7.962

Table A.17: Operation costs of ethylene production based on carbon monoxide and hydrogen feedstock on 2005.

labor					
operating	4.4	$/hr	7	operators/shift	0.084
maintenance	0.03	of total unit investment			0.802
control laboratory	0.2	of operating labor			0.017
total labor					0.903
materials					
raw & process			untis/lb		¢/lb
CO	11.11	¢/lb	2.132	lb	23.687
H2	36.84	¢/lb	0.311	lb	11.457
caustic soda	16.96	¢/lb	0.01	lb	0.170
hydrogentation cat	8.77	$/lb	0.0000056	lb	0.005
desiccant	7.95	$/lb	0.000084	lb	0.067
maintenance	3%yr of process unit investment				0.802
operating	10% of operating labor				0.008
Total materials					36.196
untilities			untis/lb		¢/lb
CW	0.01	¢/gal	60.322	gal	0.603
Steam	0.32	¢/lb	5.689	lb	1.820
Electricity	5.85	¢/kwh	0.355	kw	2.077
Process water	0.18	¢/gal	0.114	gal	0.021
Fuel	0.000199	¢/btu	2097	btu	0.417
Total Utilities					4.938
Total Direct Operating costs					42.037
plant overhead	0.8	% of total labor	0.722		
taxes & insurance	0.02	% of fixed capital	0.683		
plant cost			43.443		
sales & research	0.06	% of sales	NA		
cash expenditures			43.443		
depreciation	0.1	%/yr of fixed capital	3.417		
interest on working capital	0.06	%/yr	0.139		
total product cost			46.999		
Indirect costs					4.962
by-product credit			untis/lb		¢/lb
methane	4.27	¢/lb	0.073	lb	0.312
					0.312
Net Production Cost					46.687

Table A.18: Operation costs of ethylene production based on natural gas feedstock on 2005.

labor					
operating	4.4	$/hr	7	operators/shift	0.084
maintenance	0.03	of total unit investment			0.611
control laboratory	0.2	of operating labor			0.017
total labor					0.712
materials					
raw & process			untis/lb		¢/lb
NG	1.03764228	¢/lb	1.68	lb	1.743
MEA	98	¢/lb	0.000056	lb	0.005
NMP	205	¢/lb	0.0007	lb	0.144
oxygen (95%)	0.052	¢/lb	3.83	lb	0.199
hydrogentation cat	0	$/lb	0.001986317	lb	0.000
maintenance	3%yr of process unit investment				0.611
operating	10% of operating labor				0.008
Total materials					2.711
untilities			untis/lb		¢/lb
CW	0.01	¢/gal	41.228	gal	0.412
Steam	0.32	¢/lb	5.346	lb	1.711
Electricity	5.85	¢/kwh	0.051	kw	0.298
Process water	0.18	¢/gal	1.631	gal	0.293
Total Utilities					2.714
Total Direct Operating costs					6.137
plant overhead	0.8	% of total labor	0.569		
taxes & insurance	0.02	% of fixed capital	0.562		
plant cost			7.268		
sales & research	0.06	% of sales	NA		
cash expenditures			7.268		
depreciation	0.1	%/yr of fixed capital	2.811		
interest on working capital	0.06	%/yr	0.186		
total product cost			10.265		
Indirect costs					4.128
by-product credit					
Net Production Cost					10.265

APPENDIX B

The following data are the output files of Promax software (cracking, compression, hydrogenation, and purification sections).

Recoveries Report

Component Recoveries - Project Inlets				Status:	Unsolved
Recovery Stream Data Source - All Inlets in Project					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Compressor	1	Compressor	14	Compressor	17
Compressor	18	Compressor	19	Cracker	C O
Cracker	CH4	Cracker	CO	Cracker	COKE
Cracker	CO2	Cracker	C2H2	Cracker	C2H4
Cracker	C6'S	Cracker	H 2	Cracker	H2
Cracker	H2O	Cracker	O2	Cracker	Steam
Cracker	1	Cracker	13	Cracker	14
Cracker	46	Cracker	53	Cracker	55
Ethylene purification	4	Ethylene purification	14	Ethylene purification	25
Ethylene purification	27	Ethylene purification	29	Ethylene purification	31
Ethylene purification	33	Ethylene purification	35	Hydrogenation	ETHYLENE IN
Hydrogenation	H2 IN	Hydrogenation	1	Hydrogenation	9
Hydrogenation	37	Hydrogenation	39	Hydrogenation	42
Hydrogenation	43	Hydrogenation	45	Hydrogenation	48
Hydrogenation	49	Hydrogenation	51		

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	FALSE

Component Recoveries - Project Outlets				Status:	Unsolved
Recovery Stream Data Source - All Outlets in Project					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Compressor	6	Compressor	13	Compressor	15
Compressor	16	Compressor	20	Compressor	21
Cracker	COKE_OUT	Cracker	12	Cracker	15
Cracker	22	Cracker	31	Cracker	52
Cracker	54	Ethylene purification	2	Ethylene purification	6
Ethylene purification	10	Ethylene purification	12	Ethylene purification	15
Ethylene purification	20	Ethylene purification	21	Ethylene purification	23
Ethylene purification	24	Ethylene purification	26	Ethylene purification	28
Ethylene purification	30	Ethylene purification	32	Ethylene purification	34
Ethylene purification	36	Hydrogenation	H2/C2H2 OUT	Hydrogenation	28
Hydrogenation	29	Hydrogenation	33	Hydrogenation	38
Hydrogenation	40	Hydrogenation	41	Hydrogenation	44
Hydrogenation	46	Hydrogenation	47	Hydrogenation	50
Hydrogenation	52	Hydrogenation	53		

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE

Calculate Ratios	FALSE	Summation Only	FALSE

Component Recoveries - Project Losses

Status: Unsolved

Reference Stream Data Source - All Outlets in Project

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Compressor	6	Compressor	13	Compressor	15
Compressor	16	Compressor	20	Compressor	21
Cracker	COKE_OUT	Cracker	12	Cracker	15
Cracker	22	Cracker	31	Cracker	52
Cracker	54	Ethylene purification	2	Ethylene purification	6
Ethylene purification	10	Ethylene purification	12	Ethylene purification	15
Ethylene purification	20	Ethylene purification	21	Ethylene purification	23
Ethylene purification	24	Ethylene purification	26	Ethylene purification	28
Ethylene purification	30	Ethylene purification	32	Ethylene purification	34
Ethylene purification	36	Hydrogenation	H2/C2H2 OUT	Hydrogenation	28
Hydrogenation	29	Hydrogenation	33	Hydrogenation	38
Hydrogenation	40	Hydrogenation	41	Hydrogenation	44
Hydrogenation	46	Hydrogenation	47	Hydrogenation	50
Hydrogenation	52	Hydrogenation	53		

Recovery Stream Data Source - All Inlets in Project

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Compressor	1	Compressor	14	Compressor	17
Compressor	18	Compressor	19	Cracker	C O
Cracker	CH4	Cracker	CO	Cracker	COKE
Cracker	CO2	Cracker	C2H2	Cracker	C2H4
Cracker	C6'S	Cracker	H 2	Cracker	H2
Cracker	H2O	Cracker	O2	Cracker	Steam
Cracker	1	Cracker	13	Cracker	14
Cracker	46	Cracker	53	Cracker	55
Ethylene purification	4	Ethylene purification	14	Ethylene purification	25
Ethylene purification	27	Ethylene purification	29	Ethylene purification	31
Ethylene purification	33	Ethylene purification	35	Hydrogenation	ETHYLENE IN
Hydrogenation	H2 IN	Hydrogenation	1	Hydrogenation	9
Hydrogenation	37	Hydrogenation	39	Hydrogenation	42
Hydrogenation	43	Hydrogenation	45	Hydrogenation	48
Hydrogenation	49	Hydrogenation	51		

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	TRUE

Component Recoveries - Project Recoveries

Status: Unsolved

Reference Stream Data Source - All Inlets in Project

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Compressor	1	Compressor	14	Compressor	17
Compressor	18	Compressor	19	Cracker	C O
Cracker	CH4	Cracker	CO	Cracker	COKE
Cracker	CO2	Cracker	C2H2	Cracker	C2H4

Cracker	C6'S	Cracker	H 2	Cracker	H2
Cracker	H2O	Cracker	O2	Cracker	Steam
Cracker	1	Cracker	13	Cracker	14
Cracker	46	Cracker	53	Cracker	55
Ethylene purification	4	Ethylene purification	14	Ethylene purification	25
Ethylene purification	27	Ethylene purification	29	Ethylene purification	31
Ethylene purification	33	Ethylene purification	35	Hydrogenation	ETHYLENE IN
Hydrogenation	H2 IN	Hydrogenation	1	Hydrogenation	9
Hydrogenation	37	Hydrogenation	39	Hydrogenation	42
Hydrogenation	43	Hydrogenation	45	Hydrogenation	48
Hydrogenation	49	Hydrogenation	51		

Recovery Stream Data Source - All Outlets in Project

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Compressor	6	Compressor	13	Compressor	15
Compressor	16	Compressor	20	Compressor	21
Cracker	COKE_OUT	Cracker	12	Cracker	15
Cracker	22	Cracker	31	Cracker	52
Cracker	54	Ethylene purification	2	Ethylene purification	6
Ethylene purification	10	Ethylene purification	12	Ethylene purification	15
Ethylene purification	20	Ethylene purification	21	Ethylene purification	23
Ethylene purification	24	Ethylene purification	26	Ethylene purification	28
Ethylene purification	30	Ethylene purification	32	Ethylene purification	34
Ethylene purification	36	Hydrogenation	H2/C2H2 OUT	Hydrogenation	28
Hydrogenation	29	Hydrogenation	33	Hydrogenation	38
Hydrogenation	40	Hydrogenation	41	Hydrogenation	44
Hydrogenation	46	Hydrogenation	47	Hydrogenation	50
Hydrogenation	52	Hydrogenation	53		

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	TRUE	Summation Only	FALSE

Component Recoveries - Cracker Inlets — **Status: Solved**

Recovery Stream Data Source - All Inlets in Flowsheet

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Cracker	C O	Cracker	CH4	Cracker	CO
Cracker	COKE	Cracker	CO2	Cracker	C2H2
Cracker	C2H4	Cracker	C6'S	Cracker	H 2
Cracker	H2	Cracker	H2O	Cracker	O2
Cracker	Steam	Cracker	1	Cracker	13
Cracker	14	Cracker	46	Cracker	53
Cracker	55				

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	FALSE

Tabulated Data

Index	Cracker:C O lbmol/h	Cracker:CH4 lbmol/h	Cracker:CO lbmol/h	Cracker:COKE lbmol/h	Cracker:CO2 lbmol/h
Water	0	0	0	0	0
Carbon Dioxide	0	0	0	0	1368.93
Carbon Monoxide	1370.98	0	951.293	0	0
H2	0	0	0	0	0
Oxygen	0	0	0	0	0
Nitrogen	0	0	0	0	0
Methane	0	2041.03	0	0	0
Ethane	0	0	0	0	0
Ethylene	0	0	0	0	0
Acetylene	0	0	0	0	0
Propane	0	0	0	0	0
Propylene	0	0	0	0	0
n-Butane	0	0	0	0	0
i-Butane	0	0	0	0	0
n-Pentane	0	0	0	0	0
i-Pentane	0	0	0	0	0
n-Hexane	0	0	0	0	0
Naphthalene	0	0	0	20.9755	0

Index	Cracker:C2H2 lbmol/h	Cracker:C2H4 lbmol/h	Cracker:C6'S lbmol/h	Cracker:H 2 lbmol/h	Cracker:H2 lbmol/h
Water	0	0	0	0	0
Carbon Dioxide	0	0	0	0	0
Carbon Monoxide	0	0	0	0	0
H2	0	0	0	6119.10	1351.17
Oxygen	0	0	0	0	0
Nitrogen	0	0	0	0	0
Methane	0	0	0	0	0
Ethane	0	0	0	0	0
Ethylene	0	94.3898	0	0	0
Acetylene	1450.81	0	0	0	0
Propane	0	0	0	0	0
Propylene	0	0	0	0	0
n-Butane	0	0	0	0	0
i-Butane	0	0	0	0	0
n-Pentane	0	0	0	0	0
i-Pentane	0	0	0	0	0
n-Hexane	0	0	46.6122	0	0

Index	Cracker:H2O lbmol/h	Cracker:O2 lbmol/h	Cracker:Steam lbmol/h	Cracker:1 lbmol/h	Cracker:13 lbmol/h
Water	4523.48	0	1646.97	33.5917	0
Carbon Dioxide	0	0	0	6.29845	0
Carbon Monoxide	0	0	0	1149.02	0
H2	0	0	0	3730.48	0
Oxygen	0	685.492	0	0	0
Nitrogen	0	0	0	278.331	0
Methane	0	0	0	988.556	0
Ethane	0	0	0	3.29919	0
Ethylene	0	0	0	61.1849	0
Acetylene	0	0	0	21.5947	0
Propane	0	0	0	0.299926	0
Propylene	0	0	0	0.599852	0
n-Butane	0	0	0	0	0
i-Butane	0	0	0	2.39941	0

n-Pentane	0	0	0	0	0
i-Pentane	0	0	0	2.09948	0
n-Hexane	0	0	0	2.69933	0

Index	Cracker:14 lbmol/h	Cracker:46 lbmol/h	Cracker:53 lbmol/h	Cracker:55 lbmol/h	Summary Table lbmol/h
Water	0	0	548990	516419	1.07161E+06
Carbon Dioxide	0	135.918	0	7.37149	1518.52
Carbon Monoxide	0	0	0	2.13599	3473.43
H2	0	24.6033	0	2.84108	11228.2
Oxygen	4217.31	0	0	0	4902.80
Nitrogen	219.263	63.2084	0	0.304936	561.108
Methane	0	4142.05	0	0.785900	7172.43
Ethane	0	593.579	0	0	596.878
Ethylene	0	0	0	0.0906023	155.665
Acetylene	0	0	0	4.15272	1476.55
Propane	0	431.258	0	0	431.558
Propylene	0	0	0	0	0.599852
n-Butane	0	64.7086	0	0	64.7086
i-Butane	0	27.5037	0	0	29.9031
n-Pentane	0	4.40059	0	0	4.40059
i-Pentane	0	0.200027	0	0	2.29951
n-Hexane	0	2.47033	0	0.00582448	51.7877
Naphthalene	0	0	0	0	20.9755

Component Recoveries - Cracker Outlets Status: **Solved**

Recovery Stream Data Source - All Outlets in Flowsheet

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Cracker	COKE_OUT	Cracker	12	Cracker	15
Cracker	22	Cracker	31	Cracker	52
Cracker	54				

Tabulated Data

Index	Cracker:COKE_OUT lbmol/h	Cracker:12 lbmol/h	Cracker:15 lbmol/h	Cracker:22 lbmol/h	Cracker:31 lbmol/h
Water	0	1571.32	0	0	4667.09
Carbon Dioxide	0	1509.40	0	0	0.0656526
Carbon Monoxide	0	2323.26	1149.02	0	0.00645475
H2	0	7494.55	3730.48	0	0.0235573
Oxygen	0	0	4217.31	685.492	0
Nitrogen	0	560.922	0	0	0.00110316
Methane	0	2040.85	988.556	4142.05	0.00722713
Ethane	0	0	3.29919	593.579	0
Ethylene	0	94.3685	61.1849	0	0.000824147
Acetylene	0	1441.92	21.5947	0	0.110128
Propane	0	0	0.299926	431.258	0
Propylene	0	0	0.599852	0	0
n-Butane	0	0	0	64.7086	0
i-Butane	0	0	2.39941	27.5037	0
n-Pentane	0	0	0	4.40059	0

i-Pentane	0	0	2.09948	0.200027	0
n-Hexane	0	46.6108	2.69933	2.47033	5.29970E-05
Naphthalene	20.9755	0	0	0	0

Index	Cracker:52 lbmol/h	Cracker:54 lbmol/h	Summary Table lbmol/h
Water	548990	516419	1.07165E+06
Carbon Dioxide	0	7.37149	1516.84
Carbon Monoxide	0	2.13599	3474.42
H2	0	2.84108	11227.9
Oxygen	0	0	4902.80
Nitrogen	0	0.304936	561.228
Methane	0	0.785900	7172.25
Ethane	0	0	596.878
Ethylene	0	0.0906023	155.645
Acetylene	0	4.15272	1467.78
Propane	0	0	431.558
Propylene	0	0	0.599852
n-Butane	0	0	64.7086
i-Butane	0	0	29.9031
n-Pentane	0	0	4.40059
i-Pentane	0	0	2.29951
n-Hexane	0	0.00582448	51.7864
Naphthalene	0	0	20.9755

Component Recoveries - Cracker Losses Status: **Solved**

Reference Stream Data Source - All Outlets in Flowsheet

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Cracker	COKE_OUT	Cracker	12	Cracker	15
Cracker	22	Cracker	31	Cracker	52
Cracker	54				

Recovery Stream Data Source - All Inlets in Flowsheet

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Cracker	C O	Cracker	CH4	Cracker	CO
Cracker	COKE	Cracker	CO2	Cracker	C2H2
Cracker	C2H4	Cracker	C6'S	Cracker	H 2
Cracker	H2	Cracker	H2O	Cracker	O2
Cracker	Steam	Cracker	1	Cracker	13
Cracker	14	Cracker	46	Cracker	53
Cracker	55				

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	TRUE

Tabulated Data

Index	Summary Table lbmol/h
Water	-34.3667
Carbon Dioxide	1.68159
Carbon Monoxide	-0.988828
H2	0.304069
Oxygen	0
Nitrogen	-0.120078
Methane	0.177507
Ethane	-1.12786E-13
Ethylene	0.0204524
Acetylene	8.77778
Propane	-5.63932E-14
Propylene	0
n-Butane	0
i-Butane	0
n-Pentane	0
i-Pentane	0
n-Hexane	0.00132530
Naphthalene	-3.52458E-15

Component Recoveries - Cracker Recoveries Status: Solved

Reference Stream Data Source - All Inlets in Flowsheet

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Cracker	C O	Cracker	CH4	Cracker	CO
Cracker	COKE	Cracker	CO2	Cracker	C2H2
Cracker	C2H4	Cracker	C6'S	Cracker	H 2
Cracker	H2	Cracker	H2O	Cracker	O2
Cracker	Steam	Cracker	1	Cracker	13
Cracker	14	Cracker	46	Cracker	53
Cracker	55				

Recovery Stream Data Source - All Outlets in Flowsheet

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Cracker	COKE_OUT	Cracker	12	Cracker	15
Cracker	22	Cracker	31	Cracker	52
Cracker	54				

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	TRUE	Summation Only	FALSE

Tabulated Data

Index	Cracker:COKE_OUT %	Cracker:12 %	Cracker:15 %	Cracker:22 %	Cracker:31 %
Water	0	0.146632	0	0	0.435519
Carbon	0	99.3995	0	0	0.00432345

Dioxide					
Carbon Monoxide	0	66.8866	33.0802	0	0.000185832
H2	0	66.7476	33.2242	0	0.000209805
Oxygen	0	0	86.0183	13.9817	0
Nitrogen	0	99.9669	0	0	0.000196604
Methane	0	28.4540	13.7827	57.7497	0.000100763
Ethane	0	0	0.552740	99.4473	0
Ethylene	0	60.6227	39.3054	0	0.000529435
Acetylene	0	97.6543	1.46251	0	0.00745847
Propane	0	0	0.0694985	99.9305	0
Propylene	0	0	100	0	0
n-Butane	0	0	0	100	0
i-Butane	0	0	8.02395	91.9761	0
n-Pentane	0	0	0	100	0
i-Pentane	0	0	91.3013	8.69867	0
n-Hexane	0	90.0037	5.21230	4.77011	0.000102335
Naphthalene	100	0	0	0	0

Index	Cracker:52 %	Cracker:54 %	Summary Table %
Water	51.2302	48.1908	100.003
Carbon Dioxide	0	0.485439	99.8893
Carbon Monoxide	0	0.0614952	100.028
H2	0	0.0253030	99.9973
Oxygen	0	0	100
Nitrogen	0	0.0543454	100.021
Methane	0	0.0109572	99.9975
Ethane	0	0	100
Ethylene	0	0.0582033	99.9869
Acetylene	0	0.281244	99.4055
Propane	0	0	100
Propylene	0	0	100
n-Butane	0	0	100
i-Butane	0	0	100
n-Pentane	0	0	100
i-Pentane	0	0	100
n-Hexane	0	0.0112468	99.9974
Naphthalene	0	0	100

Component Recoveries - Compressor Inlets **Status:** **Solved**

Recovery Stream Data Source - All Inlets in Flowsheet

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Compressor	1	Compressor	14	Compressor	17
Compressor	18	Compressor	19		

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	FALSE

Tabulated Data

Index	Compressor:1 lbmol/h	Compressor:14 lbmol/h	Compressor:17 lbmol/h	Compressor:18 lbmol/h	Compressor:19 lbmol/h

Water	1603	737.419	21959.6	414.743	24155.6
Carbon Dioxide	1509	1508.95	0	1508.92	0
Carbon Monoxide	2320	2319.99	0	2319.98	0
H2	7519	7518.99	0	7518.98	0
Oxygen	0	0	0	0	0
Nitrogen	562	561.998	0	561.998	0
Methane	2030	2030.00	0	2029.99	0
Ethane	7.3	7.29998	0	7.29997	0
Ethylene	92.8	92.7995	0	92.7992	0
Acetylene	1448	1447.99	0	1447.98	0
Propane	1.1	1.10000	0	1.10000	0
Propylene	1.6	1.59999	0	1.59998	0
n-Butane	0	0	0	0	0
i-Butane	11.1	11.1000	0	11.1000	0
n-Pentane	0	0	0	0	0
i-Pentane	20.1	20.1000	0	20.1000	0
n-Hexane	34	34.0000	0	34.0000	0

Index	Summary Table lbmol/h
Water	48870.4
Carbon Dioxide	4526.86
Carbon Monoxide	6959.97
H2	22557.0
Oxygen	0
Nitrogen	1686.00
Methane	6089.99
Ethane	21.9000
Ethylene	278.399
Acetylene	4343.96
Propane	3.29999
Propylene	4.79997
n-Butane	0
i-Butane	33.3000
n-Pentane	0
i-Pentane	60.3000
n-Hexane	102.000

Component Recoveries - Compressor Outlets Status: Solved

Recovery Stream Data Source - All Outlets in Flowsheet

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Compressor	6	Compressor	13	Compressor	15
Compressor	16	Compressor	20	Compressor	21

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	FALSE

Tabulated Data

Index	Compressor:6 lbmol/h	Compressor:13 lbmol/h	Compressor:15 lbmol/h	Compressor:16 lbmol/h	Compressor:20 lbmol/h
Water	1487.94	115.061	737.419	21959.6	414.743
Carbon Dioxide	0.191382	1508.81	1508.95	0	1508.92

Carbon Monoxide	0.0121419	2319.99	2319.99	0	2319.98
H2	0.0366118	7518.96	7518.99	0	7518.98
Oxygen	0	0	0	0	0
Nitrogen	0.00207143	561.998	561.998	0	561.998
Methane	0.0142310	2029.99	2030.00	0	2029.99
Ethane	5.60273E-05	7.29994	7.29998	0	7.29997
Ethylene	0.00178204	92.7982	92.7995	0	92.7992
Acetylene	0.294889	1447.71	1447.99	0	1447.98
Propane	7.73746E-06	1.09999	1.10000	0	1.10000
Propylene	4.00730E-05	1.59996	1.59999	0	1.59998
n-Butane	0	0	0	0	0
i-Butane	2.97316E-05	11.1000	11.1000	0	11.1000
n-Pentane	0	0	0	0	0
i-Pentane	5.95933E-05	20.0999	20.1000	0	20.1000
n-Hexane	0.000100446	33.9999	34.0000	0	34.0000

Index	**Compressor:21 lbmol/h**	**Summary Table lbmol/h**
Water	24155.6	48870.4
Carbon Dioxide	0	4526.86
Carbon Monoxide	0	6959.97
H2	0	22557.0
Oxygen	0	0
Nitrogen	0	1686.00
Methane	0	6089.99
Ethane	0	21.9000
Ethylene	0	278.399
Acetylene	0	4343.96
Propane	0	3.29999
Propylene	0	4.79997
n-Butane	0	0
i-Butane	0	33.3000
n-Pentane	0	0
i-Pentane	0	60.3000
n-Hexane	0	102.000

Component Recoveries - Compressor Losses **Status:** **Solved**

Reference Stream Data Source - All Outlets in Flowsheet

Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Compressor	6	Compressor	13	Compressor	15
Compressor	16	Compressor	20	Compressor	21

Recovery Stream Data Source - All Inlets in Flowsheet

Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Compressor	1	Compressor	14	Compressor	17
Compressor	18	Compressor	19		

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	TRUE

Tabulated Data

Index	**Summary Table lbmol/h**

Water	1.60563E-05
Carbon Dioxide	-1.55674E-06
Carbon Monoxide	-2.39502E-06
H2	-7.76216E-06
Oxygen	0
Nitrogen	-5.80179E-07
Methane	-2.09562E-06
Ethane	-7.53596E-09
Ethylene	-9.57933E-08
Acetylene	-1.49315E-06
Propane	-1.13556E-09
Propylene	-1.65155E-09
n-Butane	0
i-Butane	-1.14591E-08
n-Pentane	0
i-Pentane	-2.07503E-08
n-Hexane	-3.50999E-08

Component Recoveries - Compressor Recoveries				Status:	Solved
Reference Stream Data Source - All Inlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Compressor	1	Compressor	14	Compressor	17
Compressor	18	Compressor	19		
Recovery Stream Data Source - All Outlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Compressor	6	Compressor	13	Compressor	15
Compressor	16	Compressor	20	Compressor	21

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	TRUE	Summation Only	FALSE

Tabulated Data

Index	Compressor:6 %	Compressor:13 %	Compressor:15 %	Compressor:16 %	Compressor:20 %
Water	3.04467	0.235441	1.50893	44.9344	0.848660
Carbon Dioxide	0.00422770	33.3301	33.3332	0	33.3325
Carbon Monoxide	0.000174454	33.3333	33.3333	0	33.3332
H2	0.000162308	33.3332	33.3333	0	33.3333
Oxygen	0	0	0	0	0
Nitrogen	0.000122861	33.3333	33.3333	0	33.3333
Methane	0.000233679	33.3332	33.3333	0	33.3333
Ethane	0.000255833	33.3331	33.3333	0	33.3333
Ethylene	0.000640103	33.3329	33.3333	0	33.3332
Acetylene	0.00678847	33.3268	33.3333	0	33.3331
Propane	0.000234469	33.3332	33.3333	0	33.3333
Propylene	0.000834860	33.3327	33.3333	0	33.3332
n-Butane	0	0	0	0	0
i-Butane	8.92841E-05	33.3333	33.3333	0	33.3333
n-Pentane	0	0	0	0	0
i-Pentane	9.88281E-05	33.3333	33.3333	0	33.3333
n-Hexane	9.84767E-05	33.3333	33.3333	0	33.3333

	Compressor:21	**Summary Table**

Index	%	%
Water	49.4279	100.0000
Carbon Dioxide	0	100.000
Carbon Monoxide	0	100.000
H2	0	100.000
Oxygen	0	0
Nitrogen	0	100.000
Methane	0	100.000
Ethane	0	100.000
Ethylene	0	100.000
Acetylene	0	100.000
Propane	0	100.000
Propylene	0	100.000
n-Butane	0	0
i-Butane	0	100.000
n-Pentane	0	0
i-Pentane	0	100.000
n-Hexane	0	100.000

Component Recoveries - Hydrogenation Inlets				Status:	Solved
Recovery Stream Data Source - All Inlets in Flowsheet					
Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Hydrogenation	ETHYLENE IN	Hydrogenation	H2 IN	Hydrogenation	1
Hydrogenation	9	Hydrogenation	37	Hydrogenation	39
Hydrogenation	42	Hydrogenation	43	Hydrogenation	45
Hydrogenation	48	Hydrogenation	49	Hydrogenation	51

Parameters

Composition			
Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	FALSE

Tabulated Data

Index	Hydrogenation: ETHYLENE IN lbmol/h	Hydrogenation: H2 IN lbmol/h	Hydrogenation:1 lbmol/h	Hydrogenation:9 lbmol/h	Hydrogenation: 37 lbmol/h
Water	0	0	109.2	295	204.659
Carbon Dioxide	0	0	1509.4	0	18.0078
Carbon Monoxide	0	0	2320.4	0	0.612277
H2	0	476.651	7518.7	0	0.270927
Oxygen	0	0	0	0	0
Nitrogen	0	0	562.2	0	0.135911
Methane	0	0	2029.8	0	2.75333
Ethane	0	0	7.3	0	0.0959872
Ethylene	1350.68	0	92.8	0	30.1957
Acetylene	0	0	1447.9	0	0
Propane	0	0	1.1	0	0.0681283
Propylene	0	0	1.6	0	0.0985162
n-Butane	0	0	0	0	0
i-Butane	0	0	11.1	0	1.70984
n-Pentane	0	0	0	0	0
i-Pentane	0	0	20.1	0	8.91172
n-Hexane	0	0	34	0	39.0909
Naphthalene	0	0	0	0	0
NMP	0	0	0	5.3	19420.3

Index	Hydrogenation: 39 lbmol/h	Hydrogenation: 42 lbmol/h	Hydrogenation:43 lbmol/h	Hydrogenation:45 lbmol/h	Hydrogenation: 48 lbmol/h
Water	54899.0	6587.88	112.836	281.918	103.467
Carbon Dioxide	0	0	562.376	217.713	561.961
Carbon Monoxide	0	0	571.292	14.9922	571.276
H2	0	0	491.386	22.3282	491.380
Oxygen	0	0	0	0	0
Nitrogen	0	0	138.197	3.40113	138.193
Methane	0	0	529.609	45.1581	529.557
Ethane	0	0	2.60909	0.951437	2.60745
Ethylene	0	0	1336.06	8.24377	1335.49
Acetylene	0	0	0	987.669	0
Propane	0	0	0.621671	0.424612	0.620483
Propylene	0	0	0.904179	0.616942	0.902426
n-Butane	0	0	0	0	0
i-Butane	0	0	8.67144	7.68718	8.64134
n-Pentane	0	0	0	0	0
i-Pentane	0	0	17.8146	21.7077	17.6200
n-Hexane	0	0	32.5873	62.3356	31.4047
Naphthalene	0	0	0	0	0
NMP	0	0	189.589	19424.5	4.20739

Index	Hydrogenation:4! lbmol/h	Hydrogenation:5: lbmol/h	Summary Table lbmol/h
Water	1097.98	12077.8	75769.8
Carbon Dioxide	0	0	2869.46
Carbon Monoxide	0	0	3478.57
H2	0	0	9000.72
Oxygen	0	0	0
Nitrogen	0	0	842.127
Methane	0	0	3136.88
Ethane	0	0	13.5640
Ethylene	0	0	4153.47
Acetylene	0	0	2435.57
Propane	0	0	2.83489
Propylene	0	0	4.12206
n-Butane	0	0	0
i-Butane	0	0	37.8098
n-Pentane	0	0	0
i-Pentane	0	0	86.1540
n-Hexane	0	0	199.419
Naphthalene	0	0	0
NMP	0	0	39043.9

Component Recoveries - Hydrogenation Outlets				Status:	Solved
Recovery Stream Data Source - All Outlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Hydrogenation	H2/C2H2 OUT	Hydrogenation	28	Hydrogenation	29
Hydrogenation	33	Hydrogenation	38	Hydrogenation	40
Hydrogenation	41	Hydrogenation	44	Hydrogenation	46
Hydrogenation	47	Hydrogenation	50	Hydrogenation	52
Hydrogenation	53	Hydrogenation	59		

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	FALSE

Tabulated Data

Index	Hydrogenation:H2/C2H lbmol/h	Hydrogenation:28 lbmol/h	Hydrogenation:2 lbmol/h	Hydrogenation:33 lbmol/h	Hydrogenation:38 lbmol/h
Water	0	0	0	235.299	204.659
Carbon Dioxide	0	0	946.390	0.0735919	18.0078
Carbon Monoxide	0	0	0	0.00519939	0.612277
H2	1827.33	0	0	0.0151002	0.270927
Oxygen	0	0	0	0	0
Nitrogen	0	0	0	0.000888545	0.135911
Methane	0	0	0	0.00602262	2.75333
Ethane	0	0	0	2.11473E-05	0.0959872
Ethylene	0	0	0	0.00102316	30.1957
Acetylene	1350.68	0	0	0.0118905	0
Propane	0	0	0	2.05674E-06	0.0681283
Propylene	0	0	0	1.01369E-05	0.0985162
n-Butane	0	0	0	0	0
i-Butane	0	0	0	3.50255E-06	1.70984
n-Pentane	0	0	0	0	0
i-Pentane	0	0	0	4.57976E-06	8.91172
n-Hexane	0	0	0	5.45400E-06	39.0909
Naphthalene	0	0	0	0	0
NMP	0	0	0	0.000297314	19420.3

Index	Hydrogenation:4 lbmol/h	Hydrogenation:41 lbmol/h	Hydrogenation:4 lbmol/h	Hydrogenation:46 lbmol/h	Hydrogenation:47 lbmol/h
Water	54899.0	6587.88	1097.98	112.836	281.918
Carbon Dioxide	0	0	0	562.376	217.713
Carbon Monoxide	0	0	0	571.292	14.9922
H2	0	0	0	491.386	22.3282
Oxygen	0	0	0	0	0
Nitrogen	0	0	0	138.197	3.40113
Methane	0	0	0	529.609	45.1581
Ethane	0	0	0	2.60909	0.951437
Ethylene	0	0	0	1336.06	8.24377
Acetylene	0	0	0	0	987.669
Propane	0	0	0	0.621671	0.424612
Propylene	0	0	0	0.904179	0.616942
n-Butane	0	0	0	0	0
i-Butane	0	0	0	8.67144	7.68718
n-Pentane	0	0	0	0	0
i-Pentane	0	0	0	17.8146	21.7077
n-Hexane	0	0	0	32.5873	62.3356
Naphthalene	0	0	0	0	0
NMP	0	0	0	189.589	19424.5

Index	Hydrogenation:5 lbmol/h	Hydrogenation:52 lbmol/h	Hydrogenation:5 lbmol/h	Hydrogenation:59 lbmol/h	Summary Table lbmol/h
Water	103.467	12077.8	67.5558	104.741	75773.2
Carbon Dioxide	561.961	0	0	562.808	2869.33
Carbon Monoxide	571.276	0	1745.13	574.929	3478.24
H2	491.380	0	5691.66	476.391	9000.76
Oxygen	0	0	0	0	0
Nitrogen	138.193	0	423.234	138.886	842.048
Methane	529.557	0	1498.89	530.642	3136.61
Ethane	2.60745	0	4.67608	2.61914	13.5592
Ethylene	1335.49	0	91.7077	1351.66	4153.36
Acetylene	0	0	97.2079	0	2435.57
Propane	0.620483	0	0.465664	0.633896	2.83446
Propylene	0.902426	0	0.679196	0.921704	4.12297
n-Butane	0	0	0	0	0
i-Butane	8.64134	0	2.13800	8.93423	37.7820
n-Pentane	0	0	0	0	0
i-Pentane	17.6200	0	2.43172	17.7491	86.2348
n-Hexane	31.4047	0	3.01920	31.1664	199.604
Naphthalene	0	0	0	0	0
NMP	4.20739	0	7.10176	4.20412	39049.9

Component Recoveries - Hydrogenation Losses				Status:	Solved
Reference Stream Data Source - All Outlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Hydrogenation	H2/C2H2 OUT	Hydrogenation	28	Hydrogenation	29
Hydrogenation	33	Hydrogenation	38	Hydrogenation	40
Hydrogenation	41	Hydrogenation	44	Hydrogenation	46
Hydrogenation	47	Hydrogenation	50	Hydrogenation	52
Hydrogenation	53	Hydrogenation	59		
Recovery Stream Data Source - All Inlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Hydrogenation	ETHYLENE IN	Hydrogenation	H2 IN	Hydrogenation	1
Hydrogenation	9	Hydrogenation	37	Hydrogenation	39
Hydrogenation	42	Hydrogenation	43	Hydrogenation	45
Hydrogenation	48	Hydrogenation	49	Hydrogenation	51

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	TRUE

Tabulated Data

Index	Summary Table lbmol/h
Water	-3.39575
Carbon Dioxide	0.128493
Carbon Monoxide	0.334511
H2	-0.0427391
Oxygen	0
Nitrogen	0.0786384
Methane	0.264049
Ethane	0.00475079
Ethylene	0.110447
Acetylene	0
Propane	0.000437471
Propylene	-0.000910596
n-Butane	0
i-Butane	0.0277640
n-Pentane	0
i-Pentane	-0.0807979
n-Hexane	-0.185622
Naphthalene	0
NMP	-6.00617

Component Recoveries - Hydrogenation Recoveries				Status:	Solved
Reference Stream Data Source - All Inlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Hydrogenation	ETHYLENE IN	Hydrogenation	H2 IN	Hydrogenation	1
Hydrogenation	9	Hydrogenation	37	Hydrogenation	39
Hydrogenation	42	Hydrogenation	43	Hydrogenation	45
Hydrogenation	48	Hydrogenation	49	Hydrogenation	51
Recovery Stream Data Source - All Outlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Hydrogenation	H2/C2H2 OUT	Hydrogenation	28	Hydrogenation	29
Hydrogenation	33	Hydrogenation	38	Hydrogenation	40
Hydrogenation	41	Hydrogenation	44	Hydrogenation	46
Hydrogenation	47	Hydrogenation	50	Hydrogenation	52
Hydrogenation	53	Hydrogenation	59		
Parameters					
Composition Basis	Molar Flow*		Atomic Basis	FALSE	
Calculate Ratios	TRUE		Summation Only	FALSE	

Tabulated Data

Index	Hydrogenation:H2/C2H %	Hydrogenation:28 %	Hydrogenation:2 %	Hydrogenation:33 %	Hydrogenation:38 %
Water	0	0	0	0.310545	0.270106
Carbon Dioxide	0	0	32.9815	0.00256466	0.627568
Carbon Monoxide	0	0	0	0.000149469	0.0176014
H2	20.3021	0	0	0.000167766	0.00301006
Oxygen	0	0	0	0	0
Nitrogen	0	0	0	0.000105512	0.0161390
Methane	0	0	0	0.000191994	0.0877729
Ethane	0	0	0	0.000155908	0.707663
Ethylene	0	0	0	2.46339E-05	0.727001
Acetylene	55.4565	0	0	0.000488204	0
Propane	0	0	0	7.25510E-05	2.40320
Propylene	0	0	0	0.000245919	2.38997
n-Butane	0	0	0	0	0
i-Butane	0	0	0	9.26361E-06	4.52222
n-Pentane	0	0	0	0	0
i-Pentane	0	0	0	5.31578E-06	10.3439
n-Hexane	0	0	0	2.73495E-06	19.6025
Naphthalene	0	0	0	0	0
NMP	0	0	0	7.61487E-07	49.7396

Index	Hydrogenation:4 %	Hydrogenation:41 %	Hydrogenation:4 %	Hydrogenation:46 %	Hydrogenation:47 %
Water	72.4551	8.69461	1.44910	0.148919	0.372072
Carbon Dioxide	0	0	0	19.5987	7.58726
Carbon Monoxide	0	0	0	16.4232	0.430988
H2	0	0	0	5.45941	0.248071
Oxygen	0	0	0	0	0
Nitrogen	0	0	0	16.4104	0.403873
Methane	0	0	0	16.8833	1.43959
Ethane	0	0	0	19.2355	7.01445
Ethylene	0	0	0	32.1674	0.198479
Acetylene	0	0	0	0	40.5519
Propane	0	0	0	21.9293	14.9780
Propylene	0	0	0	21.9351	14.9668
n-Butane	0	0	0	0	0
i-Butane	0	0	0	22.9344	20.3312
n-Pentane	0	0	0	0	0
i-Pentane	0	0	0	20.6777	25.1964
n-Hexane	0	0	0	16.3412	31.2587
Naphthalene	0	0	0	0	0
NMP	0	0	0	0.485580	49.7504

Index	Hydrogenation:5 %	Hydrogenation:52 %	Hydrogenation:5 %	Hydrogenation:59 %	Summary Table %
Water	0.136555	15.9401	0.0891593	0.138235	100.004
Carbon Dioxide	19.5842	0	0	19.6137	99.9955
Carbon Monoxide	16.4227	0	50.1680	16.5277	99.9904
H2	5.45935	0	63.2356	5.29281	100.000
Oxygen	0	0	0	0	0
Nitrogen	16.4100	0	50.2578	16.4923	99.9907
Methane	16.8817	0	47.7828	16.9163	99.9916
Ethane	19.2234	0	34.4743	19.3096	99.9650
Ethylene	32.1535	0	2.20798	32.5430	99.9973
Acetylene	0	0	3.99118	0	100
Propane	21.8874	0	16.4262	22.3605	99.9846
Propylene	21.8926	0	16.4771	22.3603	100.022
n-Butane	0	0	0	0	0
i-Butane	22.8548	0	5.65463	23.6294	99.9266
n-Pentane	0	0	0	0	0
i-Pentane	20.4517	0	2.82252	20.6016	100.094
n-Hexane	15.7481	0	1.51400	15.6286	100.093
Naphthalene	0	0	0	0	0
NMP	0.0107761	0	0.0181892	0.0107677	100.015

Component Recoveries - Ethylene purification Inlets				Status:	Unsolved
Recovery Stream Data Source - All Inlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Ethylene purification	1	Ethylene purification	4	Ethylene purification	14
Ethylene purification	25	Ethylene purification	27	Ethylene purification	29
Ethylene purification	31	Ethylene purification	33	Ethylene purification	35
Parameters					
Composition Basis	Molar Flow*		Atomic Basis	FALSE	
Calculate Ratios	FALSE		Summation Only	FALSE	

Component Recoveries - Ethylene purification Outlets				Status:	Unsolved
Recovery Stream Data Source - All Outlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Ethylene purification	2	Ethylene purification	6	Ethylene purification	10
Ethylene purification	12	Ethylene purification	15	Ethylene purification	20
Ethylene purification	21	Ethylene purification	23	Ethylene purification	24
Ethylene purification	26	Ethylene purification	28	Ethylene purification	30
Ethylene purification	32	Ethylene purification	34	Ethylene purification	36
Parameters					
Composition Basis	Molar Flow*		Atomic Basis	FALSE	
Calculate Ratios	FALSE		Summation Only	FALSE	

Component Recoveries - Ethylene purification Losses				Status:	Unsolved
Reference Stream Data Source - All Outlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Ethylene purification	2	Ethylene purification	6	Ethylene purification	10
Ethylene purification	12	Ethylene purification	15	Ethylene purification	20
Ethylene purification	21	Ethylene purification	23	Ethylene purification	24
Ethylene purification	26	Ethylene purification	28	Ethylene purification	30
Ethylene purification	32	Ethylene purification	34	Ethylene purification	36
Recovery Stream Data Source - All Inlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Ethylene purification	1	Ethylene purification	4	Ethylene purification	14
Ethylene purification	25	Ethylene purification	27	Ethylene purification	29
Ethylene purification	31	Ethylene purification	33	Ethylene purification	35
Parameters					
Composition Basis	Molar Flow*		Atomic Basis	FALSE	
Calculate Ratios	FALSE		Summation Only	TRUE	

Component Recoveries - Ethylene purification Recoveries				Status:	Unsolved
Reference Stream Data Source - All Inlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Ethylene purification	1	Ethylene purification	4	Ethylene purification	14
Ethylene purification	25	Ethylene purification	27	Ethylene purification	29
Ethylene purification	31	Ethylene purification	33	Ethylene purification	35
Recovery Stream Data Source - All Outlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Ethylene purification	2	Ethylene purification	6	Ethylene purification	10
Ethylene purification	12	Ethylene purification	15	Ethylene purification	20
Ethylene purification	21	Ethylene purification	23	Ethylene purification	24
Ethylene purification	26	Ethylene purification	28	Ethylene purification	30
Ethylene purification	32	Ethylene purification	34	Ethylene purification	36
Parameters					
Composition Basis	Molar Flow*		Atomic Basis	FALSE	
Calculate Ratios	TRUE		Summation Only	FALSE	

Environments Report

Project-Wide Constants

Atmospheric Pressure	14.6959	psia	IG Ref Volume	379.485	ft^3/lbmol
IG Ref Pressure	14.6959	psia	Liq Ref Temperature	60	°F
IG Ref Temperature	60	°F			

Environment1

Environment Settings

Number of Poynting Intervals	0		Freeze Out Temperature Threshold Difference	10	°F
Gibbs Excess Model Evaluation Temperature	77	°F	Phase Tolerance	1	%

Components

Component	Henry's Law Comp.	Phase Initiator	Component	Henry's Law Comp.	Phase Initiator
Water	FALSE	TRUE	Carbon Dioxide	FALSE	FALSE
Carbon Monoxide	FALSE	FALSE	H2	FALSE	FALSE
Oxygen	FALSE	FALSE	Nitrogen	FALSE	FALSE
Methane	FALSE	FALSE	Ethane	FALSE	FALSE
Ethylene	FALSE	FALSE	Acetylene	FALSE	FALSE
Propane	FALSE	FALSE	Propylene	FALSE	FALSE
n-Butane	FALSE	FALSE	i-Butane	FALSE	FALSE
n-Pentane	FALSE	FALSE	i-Pentane	FALSE	FALSE
n-Hexane	FALSE	FALSE	Naphthalene	FALSE	FALSE
NMP	FALSE	TRUE	Isobutylene	FALSE	FALSE
2-Methylpentane	FALSE	FALSE	3-Methylenepentane	FALSE	FALSE
Methylcyclopentane	FALSE	FALSE	Benzene	FALSE	FALSE
2,3-Dimethylpentane	FALSE	FALSE	Pentane, 2,3,4-Trimethyl-	FALSE	FALSE
Toluene	FALSE	FALSE	2,2-Dimethylhexane	FALSE	FALSE
2-Methyl-3-Ethylpentane	FALSE	FALSE	Ethylbenzene	FALSE	FALSE
m-Xylene	FALSE	FALSE	1,2,4-Trimethylbenzene	FALSE	FALSE
3-Ethyloctane	FALSE	FALSE	3-Ethylnonane	FALSE	FALSE
2,3-Dimethyldecane	FALSE	FALSE	2,3-Dimethylundecane	FALSE	FALSE
Freon-12	FALSE	FALSE			

Physical Property Method Sets

Liquid Molar Volume	COSTALD	Vapor Package	SRK
Overall Package	SRK	Light Liquid Package	SRK
Stability Calculation	SRK	Heavy Liquid Package	SRK

Environment2

Environment Settings

Number of Poynting Intervals	0		Phase Tolerance	1	%
Gibbs Excess Model Evaluation Temperature	77	°F			

Components

Component	Henry's Law Comp.	Phase Initiator	Component	Henry's Law Comp.	Phase Initiator
Water	FALSE	TRUE	Carbon Dioxide	TRUE	FALSE

Carbon Monoxide	TRUE	FALSE	H2	TRUE	FALSE
Oxygen	TRUE	FALSE	Nitrogen	TRUE	FALSE
Methane	TRUE	FALSE	Ethane	TRUE	FALSE
Ethylene	TRUE	FALSE	Acetylene	TRUE	FALSE
Propane	FALSE	FALSE	Propylene	FALSE	FALSE
n-Butane	FALSE	FALSE	i-Butane	FALSE	FALSE
n-Pentane	FALSE	FALSE	i-Pentane	FALSE	FALSE
n-Hexane	FALSE	FALSE	Naphthalene	FALSE	FALSE
NMP	FALSE	FALSE	Isobutylene	FALSE	FALSE
2-Methylpentane	FALSE	FALSE	3-Methylpentane	FALSE	FALSE
Methylcyclopentane	FALSE	FALSE	Benzene	FALSE	FALSE
2,3-Dimethylpentane	FALSE	FALSE	Pentane, 2,3,4-Trimethyl-	FALSE	FALSE
Toluene	FALSE	FALSE	2,2-Dimethylhexane	FALSE	FALSE
2-Methyl-3-Ethylpentane	FALSE	FALSE	Ethylbenzene	FALSE	FALSE
m-Xylene	FALSE	FALSE	1,2,4-Trimethylbenzene	FALSE	FALSE
3-Ethyloctane	FALSE	FALSE	3-Ethylnonane	FALSE	FALSE
2,3-Dimethyldecane	FALSE	FALSE	2,3-Dimethylundecane	FALSE	FALSE
Freon-12	FALSE	FALSE	MEA	FALSE	TRUE

Electrolytic Reactions

Dissociation of Water	H2O = H+ + HO-	MEAH+ Proton Transfer	C2H8NO+ = C2H7NO + H+
First Dissociation of CO2	CO2 + H2O = (CHO3)- + H+	Second Dissociation of CO2	(CHO3)- = H+ + (CO3)2-
MEACOO- Hydrolysis	(C3H6NO3)- + H2O = (CHO3)- + C2H7NO		

Physical Property Method Sets

Liquid Molar Volume	COSTALD	Vapor Package	SRK
Overall Package	Electrolytic ELR - SRK	Light Liquid Package	Electrolytic ELR
Stability Calculation	SRK	Heavy Liquid Package	Electrolytic ELR

Environment3

Environment Settings

Number of Poynting Intervals	0		Freeze Out Temperature Threshold Difference	10	°F
Gibbs Excess Model Evaluation Temperature	77	°F	Phase Tolerance	1	%

Physical Property Method Sets

Liquid Molar Volume	COSTALD	Vapor Package	Peng-Robinson
Overall Package	Peng-Robinson	Light Liquid Package	Peng-Robinson
Stability Calculation	Peng-Robinson	Heavy Liquid Package	Peng-Robinson

Environment4

Environment Settings

Number of Poynting Intervals	0		Phase Tolerance	1	%
Gibbs Excess Model Evaluation Temperature	77	°F			

Components

Component	Henry's Law Comp.	Phase Initiator	Component	Henry's Law Comp.	Phase Initiator
H2	TRUE	FALSE	Helium-3	TRUE	FALSE

Physical Property Method Sets

Liquid Molar Volume	COSTALD	Vapor Package	SRK
Overall Package	Electrolytic ELR - SRK	Light Liquid Package	Electrolytic ELR
Stability Calculation	SRK	Heavy Liquid Package	Electrolytic ELR

Energy Budgets Report

Power Budget - Project Power Budget

Status: Solved

Power Budget Data Source - All in Project

Flowsheet	Block	Flowsheet	Block	Flowsheet	Block
Compressor	CG 1ST STG	Compressor	CG 2ND STG	Cracker	PUMP-100
Ethylene purification	PUMP-100	Hydrogenation	ABS FEED	Hydrogenation	BOOSTER
Hydrogenation	CMPR-100	Hydrogenation	PUMP-100	Hydrogenation	REACTOR FEED

Parameters

Net Power	32977.0	hp	Total Power Required	33338.9	hp
Total Power Supplied	361.855	hp			

Tabulated Data

Index	Block Power hp	Compression Ratio	Delta P psi	Head ft
Compressor:CG 1ST STG	4423.06	1.53090	15.5	25425.5
Compressor:CG 2ND STG	16193.1	3.94993	123	91814.6
Cracker:PUMP-100	452.111	2.52316	30	71.5169
Ethylene purification:PUMP-100	285.344	8.02706	293	704.576
Hydrogenation:ABS FEED	684.842	14.9526	219	525.656
Hydrogenation:BOOSTER	1619.60	1.92995	142	40203.5
Hydrogenation:CMPR-100	9221.98	18.8486	298	132676
Hydrogenation:PUMP-100	-361.855	0.884345	-15	-9182.77
Hydrogenation:REACTOR FEED	458.810	2.03666	150	344.875

Heat Budget - Project Duty Budget

Status: Solved

Heat Budget Data Source - All in Project

Flowsheet	Block	Flowsheet	Block	Flowsheet	Block
Compressor	CG DISCH COOLER	Compressor	CG INTERCOOLER	Cracker	CG PREHEAT
Cracker	DVDR-1	Cracker	DVDR-2	Cracker	DVDR-3
Cracker	OXYGEN PREHEAT	Cracker	SPRAY TOWER CLR	Cracker	XCHG-100
Cracker	XCHG-101	Cracker	XCHG-103	Ethylene purification	DVDR-1
Ethylene purification	DVDR-2	Ethylene purification	FAXR-100	Ethylene purification	VSSL-100
Hydrogenation	DVDR-1	Hydrogenation	DVDR-2	Hydrogenation	FAXR-100
Hydrogenation	FLASH GAS CLR	Hydrogenation	HYDROG EXOTHERM	Hydrogenation	SOLVENT CLR
Hydrogenation	SOLVENT PREHEAT				

Parameters

Net Duty	-2.28947E+08	Btu/h	Total Duty Required	8.98233E+08	Btu/h
Total Duty Supplied	1.12718E+09	Btu/h			

Tabulated Data

Index	Block Duty Btu/h	Block Highest Temperature °F	Block Lowest Temperature °F
Compressor:CG DISCH	-	416.575	104

COOLER	4.71128E+07		
Compressor:CG INTERCOOLER	- 1.81976E+07	197.475	104
Cracker:CG PREHEAT	6.33460E+07	900	80
Cracker:DVDR-1	160.469	899.988	899.988
Cracker:DVDR-2	10490.0	3423.48	3423.48
Cracker:DVDR-3	414652	356.369	356.369
Cracker:OXYGEN PREHEAT	2.74838E+07	900	68.3069
Cracker:SPRAY TOWER CLR	- 6.51880E+08	188.108	116
Cracker:XCHG-100	3.95845E+07	900	107
Cracker:XCHG-101	6.67369E+08	7092.40	898.757
Cracker:XCHG-103	- 1.80954E+08	3353.32	2677.09
Ethylene purification:DVDR-1	-44414.1	122.648	122.648
Ethylene purification:DVDR-2	326685	120.140	120.140
Ethylene purification:FAXR-100	- 6.82849E+07	235.922	120
Ethylene purification:VSSL-100	- 2.11208E+07	244.737	120.000
Hydrogenation:DVDR-1	- 5.90992E+06	202.694	202.694
Hydrogenation:DVDR-2	211819	94.3366	94.3366
Hydrogenation:FAXR-100	- 2.50847E+07	657.401	120
Hydrogenation:FLASH GAS CLR	- 1.32968E+07	300.843	104
Hydrogenation:HYDROG EXOTHERM	9.94857E+07	307.677	201.387
Hydrogenation:SOLVENT CLR	- 9.52930E+07	225.913	104
Hydrogenation:SOLVENT PREHEAT	1.000000	202.102	202

Power Budget - Cracker Power Budget

Status: **Solved**

Power Budget Data Source - All in Flowsheet

Flowsheet	Block	Flowsheet	Block	Flowsheet	Block
Cracker	PUMP-100				

Parameters

Net Power	452.111	hp	Total Power Required	452.111	hp
Total Power Supplied	0	hp			

Tabulated Data

Index	Block Power hp	Compression Ratio	Delta P psi	Head ft
Cracker:PUMP-100	452.111	2.52316	30	71.5169

Heat Budget - Cracker Duty Budget

Status: **Solved**

Heat Budget Data Source - All in Flowsheet

Flowsheet	Block	Flowsheet	Block	Flowsheet	Block
Cracker	CG PREHEAT	Cracker	DVDR-1	Cracker	DVDR-2
Cracker	DVDR-3	Cracker	OXYGEN PREHEAT	Cracker	SPRAY TOWER CLR
Cracker	XCHG-100	Cracker	XCHG-101	Cracker	XCHG-103

Parameters

Net Duty	-	Btu/h	Total Duty	7.98209E+08	Btu/h

	3.46257E+07		Required		
Total Duty Supplied	8.32834E+08	Btu/h			
Tabulated Data					
Index	**Block Duty Btu/h**	**Block Highest Temperature °F**	**Block Lowest Temperature °F**		
Cracker:CG PREHEAT	6.33460E+07	900	80		
Cracker:DVDR-1	160.469	899.988	899.988		
Cracker:DVDR-2	10490.0	3423.48	3423.48		
Cracker:DVDR-3	414652	356.369	356.369		
Cracker:OXYGEN PREHEAT	2.74838E+07	900	68.3069		
Cracker:SPRAY TOWER CLR	-6.51880E+08	188.108	116		
Cracker:XCHG-100	3.95845E+07	900	107		
Cracker:XCHG-101	6.67369E+08	7092.40	898.757		
Cracker:XCHG-103	-1.80954E+08	3353.32	2677.09		
Power Budget - Compressor Power Budget				**Status:**	**Solved**
Power Budget Data Source - All in Flowsheet					
Flowsheet	**Block**	**Flowsheet**	**Block**	**Flowsheet**	**Block**
Compressor	CG 1ST STG	Compressor	CG 2ND STG		
Parameters					
Net Power	20616.2	hp	Total Power Required	20616.2	hp
Total Power Supplied	0	hp			
Tabulated Data					
Index	**Block Power hp**	**Compression Ratio**	**Delta P psi**	**Head ft**	
Compressor:CG 1ST STG	4423.06	1.53090	15.5	25425.5	
Compressor:CG 2ND STG	16193.1	3.94993	123	91814.6	
Heat Budget - Compressor Duty Budget				**Status:**	**Solved**
Heat Budget Data Source - All in Flowsheet					
Flowsheet	**Block**	**Flowsheet**	**Block**	**Flowsheet**	**Block**
Compressor	CG DISCH COOLER	Compressor	CG INTERCOOLER		
Parameters					
Net Duty	-6.53105E+07	Btu/h	Total Duty Required	0	Btu/h
Total Duty Supplied	6.53105E+07	Btu/h			
Tabulated Data					
Index	**Block Duty Btu/h**	**Block Highest Temperature °F**	**Block Lowest Temperature °F**		
Compressor:CG DISCH COOLER	-4.71128E+07	416.575	104		
Compressor:CG INTERCOOLER	-1.81976E+07	197.475	104		
Power Budget - Hydrogenation Power Budget				**Status:**	**Solved**
Power Budget Data Source - All in Flowsheet					
Flowsheet	**Block**	**Flowsheet**	**Block**	**Flowsheet**	**Block**
Hydrogenation	ABS FEED	Hydrogenation	BOOSTER	Hydrogenation	CMPR-100
Hydrogenation	PUMP-100	Hydrogenation	REACTOR FEED		

Parameters

Net Power	11623.4	hp	Total Power Required	11985.2	hp
Total Power Supplied	361.855	hp			

Tabulated Data

Index	Block Power hp	Compression Ratio	Delta P psi	Head ft
Hydrogenation:ABS FEED	684.842	14.9526	219	525.656
Hydrogenation:BOOSTER	1619.60	1.92995	142	40203.5
Hydrogenation:CMPR-100	9221.98	18.8486	298	132676
Hydrogenation:PUMP-100	-361.855	0.884345	-15	-9182.77
Hydrogenation:REACTOR FEED	458.810	2.03666	150	344.875

Heat Budget - Hydrogenation Duty Budget

Status: Solved

Heat Budget Data Source - All in Flowsheet

Flowsheet	Block	Flowsheet	Block	Flowsheet	Block
Hydrogenation	DVDR-1	Hydrogenation	DVDR-2	Hydrogenation	FAXR-100
Hydrogenation	FLASH GAS CLR	Hydrogenation	HYDROG EXOTHERM	Hydrogenation	SOLVENT CLR
Hydrogenation	SOLVENT PREHEAT				

Parameters

Net Duty	- 3.98870E+07	Btu/h	Total Duty Required	9.96975E+07	Btu/h
Total Duty Supplied	1.39584E+08	Btu/h			

Tabulated Data

Index	Block Duty Btu/h	Block Highest Temperature °F	Block Lowest Temperature °F
Hydrogenation:DVDR-1	- 5.90992E+06	202.694	202.694
Hydrogenation:DVDR-2	211819	94.3366	94.3366
Hydrogenation:FAXR-100	- 2.50847E+07	657.401	120
Hydrogenation:FLASH GAS CLR	- 1.32968E+07	300.843	104
Hydrogenation:HYDROG EXOTHERM	9.94857E+07	307.677	201.387
Hydrogenation:SOLVENT CLR	- 9.52930E+07	225.913	104
Hydrogenation:SOLVENT PREHEAT	1.000000	202.102	202

Power Budget - Ethylene purification Power Budget

Status: Solved

Power Budget Data Source - All in Flowsheet

Flowsheet	Block	Flowsheet	Block	Flowsheet	Block
Ethylene purification	PUMP-100				

Parameters

Net Power	285.344	hp	Total Power Required	285.344	hp
Total Power Supplied	0	hp			

Tabulated Data

Index	Block Power hp	Compression Ratio	Delta P psi	Head ft

Ethylene purification:PUMP-100	285.344	8.02706	293	704.576

Heat Budget - Ethylene purification Duty Budget				Status:	Solved
Heat Budget Data Source - All in Flowsheet					
Flowsheet	**Block**	**Flowsheet**	**Block**	**Flowsheet**	**Block**
Ethylene purification	DVDR-1	Ethylene purification	DVDR-2	Ethylene purification	FAXR-100
Ethylene purification	VSSL-100				

Parameters					
Net Duty	-8.91235E+07	Btu/h	Total Duty Required	326685	Btu/h
Total Duty Supplied	8.94502E+07	Btu/h			

Tabulated Data

Index	Block Duty Btu/h	Block Highest Temperature °F	Block Lowest Temperature °F
Ethylene purification:DVDR-1	-44414.1	122.648	122.648
Ethylene purification:DVDR-2	326685	120.140	120.140
Ethylene purification:FAXR-100	-6.82849E+07	235.922	120
Ethylene purification:VSSL-100	-2.11208E+07	244.737	120.000

The following data are the output files of the last stage of ethylene purification section.

Environments Report

Project-Wide Constants

Atmospheric Pressure	14.6959	psia	IG Ref Volume	379.485	ft^3/lbmol
IG Ref Pressure	14.6959	psia	Liq Ref Temperature	60	°F
IG Ref Temperature	60	°F			

Environment1

Environment Settings

Number of Poynting Intervals	0		Freeze Out Temperature Threshold Difference	5.55556	°C
Gibbs Excess Model Evaluation Temperature	25	°C	Phase Tolerance	1	%

Components

Component	Henry's Law Comp.	Phase Initiator	Component	Henry's Law Comp.	Phase Initiator
Water	FALSE	TRUE	Carbon Dioxide	FALSE	FALSE
Carbon Monoxide	FALSE	FALSE	H2	FALSE	FALSE
N2	FALSE	FALSE	Methane	FALSE	FALSE
Ethane	FALSE	FALSE	Ethylene	FALSE	FALSE
Propane	FALSE	FALSE	Propylene	FALSE	FALSE
i-Butane	FALSE	FALSE	i-Pentane	FALSE	FALSE
n-Hexane	FALSE	FALSE	MEA	FALSE	TRUE

Physical Property Method Sets

Liquid Molar Volume	COSTALD	Vapor Package	Peng-Robinson
Overall Package	Peng-Robinson	Light Liquid Package	Peng-Robinson
Stability Calculation	Peng-Robinson	Heavy Liquid Package	Peng-Robinson

Recoveries Report

Component Recoveries - Project Inlets

Status: Solved

Recovery Stream Data Source - All Inlets in Project

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Flowsheet1	1	Flowsheet1	14	Flowsheet1	16

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	FALSE

Tabulated Data

Index	Flowsheet1:1 kmol/h	Flowsheet1:14 kmol/h	Flowsheet1:16 kmol/h	Summary Table kmol/h
Water	0	199.214	1543.91	1743.12
Carbon Dioxide	0.0316986	0	0	0.0316986
Carbon Monoxide	258.524	0	0	258.524
H2	222.292	0	0	222.292
N2	62.5773	0	0	62.5773
Methane	239.354	0	0	239.354
Ethane	1.17847	0	0	1.17847
Ethylene	600.520	0	0	600.520
Propane	0.280638	0	0	0.280638
Propylene	0.405197	0	0	0.405197
i-Butane	3.91167	0	0	3.91167
i-Pentane	7.98133	0	0	7.98133
n-Hexane	14.2332	0	0	14.2332
MEA	0.0152282	0	0	0.0152282

Component Recoveries - Project Outlets

Status: Solved

Recovery Stream Data Source - All Outlets in Project

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Flowsheet1	Ethylene Product	Flowsheet1	7	Flowsheet1	12
Flowsheet1	15	Flowsheet1	17		

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	FALSE

Tabulated Data

Index	Flowsheet1:Ethylene Product kmol/h	Flowsheet1:7 kmol/h	Flowsheet1:12 kmol/h	Flowsheet1:15 kmol/h	Flowsheet1:17 kmol/h
Water	0	0	0	199.214	1543.91
Carbon Dioxide	0.0308248	6.47250E-06	0.000867321	0	0
Carbon Monoxide	2.05911E-06	7.92371E-05	258.524	0	0
H2	1.92419E-11	3.98372E-06	222.292	0	0
N2	6.69027E-08	1.22370E-05	62.5772	0	0
Methane	0.00980522	0.00127764	239.341	0	0
Ethane	1.17053	0.00724019	0.000704200	0	0

Ethylene	594.117	0.395553	6.00699	0	0
Propane	0.0140319	0.266606	2.77194E-08	0	0
Propylene	0.0560763	0.349120	2.38162E-07	0	0
i-Butane	0.0285822	3.88308	1.50029E-08	0	0
i-Pentane	3.03491E-07	7.98132	4.77966E-17	0	0
n-Hexane	1.71794E-10	14.2332	0	0	0
MEA	2.50593E-07	0.0152279	3.77667E-13	0	0

Index	Summary Table kmol/h
Water	1743.12
Carbon Dioxide	0.0316985
Carbon Monoxide	258.524
H2	222.292
N2	62.5772
Methane	239.352
Ethane	1.17847
Ethylene	600.520
Propane	0.280638
Propylene	0.405197
i-Butane	3.91167
i-Pentane	7.98132
n-Hexane	14.2332
MEA	0.0152282

Component Recoveries - Project Losses				Status:	Solved
Reference Stream Data Source - All Outlets in Project					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Flowsheet1	Ethylene Product	Flowsheet1	7	Flowsheet1	12
Flowsheet1	15	Flowsheet1	17		
Recovery Stream Data Source - All Inlets in Project					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Flowsheet1	1	Flowsheet1	14	Flowsheet1	16
Parameters					
Composition Basis	Molar Flow*		Atomic Basis	FALSE	
Calculate Ratios	FALSE		Summation Only	TRUE	
Tabulated Data					

Index	Summary Table kmol/h
Water	0
Carbon Dioxide	1.09297E-08
Carbon Monoxide	0.000650977
H2	-8.78626E-05
N2	0.000111231
Methane	0.00153906
Ethane	-3.41951E-08
Ethylene	5.51869E-05
Propane	8.93673E-08
Propylene	6.19907E-08
i-Butane	3.24228E-06

i-Pentane	1.25376E-05
n-Hexane	3.34695E-05
MEA	2.77173E-08

Component Recoveries - Project Recoveries Status: Solved

Reference Stream Data Source - All Inlets in Project

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Flowsheet1	1	Flowsheet1	14	Flowsheet1	16

Recovery Stream Data Source - All Outlets in Project

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Flowsheet1	Ethylene Product	Flowsheet1	7	Flowsheet1	12
Flowsheet1	15	Flowsheet1	17		

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	TRUE	Summation Only	FALSE

Tabulated Data

Index	Flowsheet1:Ethylene Product %	Flowsheet1:7 %	Flowsheet1:12 %	Flowsheet1:15 %	Flowsheet1:17 %
Water	0	0	0	11.4286	88.5714
Carbon Dioxide	97.2434	0.0204189	2.73615	0	0
Carbon Monoxide	7.96486E-07	3.06498E-05	99.9997	0	0
H2	8.65610E-12	1.79211E-06	100.000	0	0
N2	1.06912E-07	1.95550E-05	99.9998	0	0
Methane	0.00409654	0.000533788	99.9947	0	0
Ethane	99.3259	0.614371	0.0597553	0	0
Ethylene	98.9338	0.0658684	1.00030	0	0
Propane	4.99999	95.0000	9.87727E-06	0	0
Propylene	13.8393	86.1607	5.87769E-05	0	0
i-Butane	0.730689	99.2692	3.83541E-07	0	0
i-Pentane	3.80251E-06	99.9998	5.98856E-16	0	0
n-Hexane	1.20699E-09	99.9998	0	0	0
MEA	0.00164559	99.9982	2.48005E-09	0	0

Index	Summary Table %
Water	100
Carbon Dioxide	100.0000
Carbon Monoxide	99.9997
H2	100.000
N2	99.9998
Methane	99.9994
Ethane	100.000
Ethylene	100.0000
Propane	100.0000
Propylene	100.0000
i-Butane	99.9999
i-Pentane	99.9998
n-Hexane	99.9998
MEA	99.9998

Component Recoveries - Flowsheet1 Inlets				Status:	Solved
Recovery Stream Data Source - All Inlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Flowsheet1	1	Flowsheet1	14	Flowsheet1	16
Parameters					
Composition Basis	Molar Flow*		Atomic Basis	FALSE	
Calculate Ratios	FALSE		Summation Only	FALSE	
Tabulated Data					
Index	**Flowsheet1:1 kmol/h**	**Flowsheet1:14 kmol/h**	**Flowsheet1:16 kmol/h**	**Summary Table kmol/h**	
Water	0	199.214	1543.91	1743.12	
Carbon Dioxide	0.0316986	0	0	0.0316986	
Carbon Monoxide	258.524	0	0	258.524	
H2	222.292	0	0	222.292	
N2	62.5773	0	0	62.5773	
Methane	239.354	0	0	239.354	
Ethane	1.17847	0	0	1.17847	
Ethylene	600.520	0	0	600.520	
Propane	0.280638	0	0	0.280638	
Propylene	0.405197	0	0	0.405197	
i-Butane	3.91167	0	0	3.91167	
i-Pentane	7.98133	0	0	7.98133	
n-Hexane	14.2332	0	0	14.2332	
MEA	0.0152282	0	0	0.0152282	

Component Recoveries - Flowsheet1 Outlets				Status:	Solved
Recovery Stream Data Source - All Outlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Flowsheet1	Ethylene Product	Flowsheet1	7	Flowsheet1	12
Flowsheet1	15	Flowsheet1	17		
Parameters					
Composition Basis	Molar Flow*		Atomic Basis	FALSE	
Calculate Ratios	FALSE		Summation Only	FALSE	
Tabulated Data					
Index	**Flowsheet1:Ethylene Product kmol/h**	**Flowsheet1:7 kmol/h**	**Flowsheet1:12 kmol/h**	**Flowsheet1:15 kmol/h**	**Flowsheet1:17 kmol/h**
Water	0	0	0	199.214	1543.91
Carbon Dioxide	0.0308248	6.47250E-06	0.000867321	0	0
Carbon Monoxide	2.05911E-06	7.92371E-05	258.524	0	0
H2	1.92419E-11	3.98372E-06	222.292	0	0
N2	6.69027E-08	1.22370E-05	62.5772	0	0
Methane	0.00980522	0.00127764	239.341	0	0
Ethane	1.17053	0.00724019	0.000704200	0	0
Ethylene	594.117	0.395553	6.00699	0	0
Propane	0.0140319	0.266606	2.77194E-08	0	0
Propylene	0.0560763	0.349120	2.38162E-07	0	0

i-Butane	0.0285822	3.88308	1.50029E-08	0	0
i-Pentane	3.03491E-07	7.98132	4.77966E-17	0	0
n-Hexane	1.71794E-10	14.2332	0	0	0
MEA	2.50593E-07	0.0152279	3.77667E-13	0	0

Index	Summary Table kmol/h
Water	1743.12
Carbon Dioxide	0.0316985
Carbon Monoxide	258.524
H2	222.292
N2	62.5772
Methane	239.352
Ethane	1.17847
Ethylene	600.520
Propane	0.280638
Propylene	0.405197
i-Butane	3.91167
i-Pentane	7.98132
n-Hexane	14.2332
MEA	0.0152282

Component Recoveries - Flowsheet1 Losses **Status:** **Solved**

Reference Stream Data Source - All Outlets in Flowsheet

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Flowsheet1	Ethylene Product	Flowsheet1	7	Flowsheet1	12
Flowsheet1	15	Flowsheet1	17		

Recovery Stream Data Source - All Inlets in Flowsheet

Flowsheet	PStream	Flowsheet	PStream	Flowsheet	PStream
Flowsheet1	1	Flowsheet1	14	Flowsheet1	16

Parameters

Composition Basis	Molar Flow*	Atomic Basis	FALSE
Calculate Ratios	FALSE	Summation Only	TRUE

Tabulated Data

Index	Summary Table kmol/h
Water	0
Carbon Dioxide	1.09297E-08
Carbon Monoxide	0.000650977
H2	-8.78626E-05
N2	0.000111231
Methane	0.00153906
Ethane	-3.41951E-08
Ethylene	5.51869E-05
Propane	8.93673E-08
Propylene	6.19907E-08
i-Butane	3.24228E-06
i-Pentane	1.25376E-05
n-Hexane	3.34695E-05
MEA	2.77173E-08

Component Recoveries - Flowsheet1 Recoveries				Status:	Solved
Reference Stream Data Source - All Inlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Flowsheet1	1	Flowsheet1	14	Flowsheet1	16
Recovery Stream Data Source - All Outlets in Flowsheet					
Flowsheet	**PStream**	**Flowsheet**	**PStream**	**Flowsheet**	**PStream**
Flowsheet1	Ethylene Product	Flowsheet1	7	Flowsheet1	12
Flowsheet1	15	Flowsheet1	17		
Parameters					
Composition Basis	Molar Flow*		Atomic Basis	FALSE	
Calculate Ratios	TRUE		Summation Only	FALSE	

Tabulated Data

Index	Flowsheet1:Ethylene Product %	Flowsheet1:7 %	Flowsheet1:12 %	Flowsheet1:15 %	Flowsheet1:17 %
Water	0	0	0	11.4286	88.5714
Carbon Dioxide	97.2434	0.0204189	2.73615	0	0
Carbon Monoxide	7.96486E-07	3.06498E-05	99.9997	0	0
H2	8.65610E-12	1.79211E-06	100.000	0	0
N2	1.06912E-07	1.95550E-05	99.9998	0	0
Methane	0.00409654	0.000533788	99.9947	0	0
Ethane	99.3259	0.614371	0.0597553	0	0
Ethylene	98.9338	0.0658684	1.00030	0	0
Propane	4.99999	95.0000	9.87727E-06	0	0
Propylene	13.8393	86.1607	5.87769E-05	0	0
i-Butane	0.730689	99.2692	3.83541E-07	0	0
i-Pentane	3.80251E-06	99.9998	5.98856E-16	0	0
n-Hexane	1.20699E-09	99.9998	0	0	0
MEA	0.00164559	99.9982	2.48005E-09	0	0

Index	Summary Table %
Water	100
Carbon Dioxide	100.0000
Carbon Monoxide	99.9997
H2	100.000
N2	99.9998
Methane	99.9994
Ethane	100.000
Ethylene	100.0000
Propane	100.0000
Propylene	100.0000
i-Butane	99.9999
i-Pentane	99.9998
n-Hexane	99.9998
MEA	99.9998

Energy Budgets Report

Power Budget - Project Power Budget

Status: Solved

Parameters

Net Power	0	kW	Total Power Required	0	kW
Total Power Supplied	0	kW			

Heat Budget - Project Duty Budget

Status: Solved

Heat Budget Data Source - All in Project

Flowsheet	Block	Flowsheet	Block	Flowsheet	Block
Flowsheet1	K-100	Flowsheet1	K-100 1	Flowsheet1	K-101
Flowsheet1	VSSL-100	Flowsheet1	VSSL-101	Flowsheet1	XCHG-100

Parameters

Net Duty	-9.64861E+07	kJ/h	Total Duty Required	8.30962E+07	kJ/h
Total Duty Supplied	1.79582E+08	kJ/h			

Tabulated Data

Index	Block Duty kJ/h	Block Highest Temperature °C	Block Lowest Temperature °C
Flowsheet1:K-100	9.59334E+06	160.196	120.061
Flowsheet1:K-100 1	-9.56522E+06	147.778	25.0868
Flowsheet1:K-101	7.35028E+07	-28.3028	-28.3256
Flowsheet1:VSSL-100	-1.53626E+07	-44.3740	-59.4672
Flowsheet1:VSSL-101	-8.01067E+07	-69.4787	-128.042
Flowsheet1:XCHG-100	-7.45477E+07	147.778	23.4357

Power Budget - Flowsheet1 Power Budget

Status: Solved

Parameters

Net Power	0	kW	Total Power Required	0	kW
Total Power Supplied	0	kW			

Heat Budget - Flowsheet1 Duty Budget

Status: Solved

Heat Budget Data Source - All in Flowsheet

Flowsheet	Block	Flowsheet	Block	Flowsheet	Block
Flowsheet1	K-100	Flowsheet1	K-100 1	Flowsheet1	K-101
Flowsheet1	VSSL-100	Flowsheet1	VSSL-101	Flowsheet1	XCHG-100

Parameters

Net Duty	-9.64861E+07	kJ/h	Total Duty Required	8.30962E+07	kJ/h
Total Duty Supplied	1.79582E+08	kJ/h			

Tabulated Data

Index	Block Duty kJ/h	Block Highest Temperature °C	Block Lowest Temperature °C
Flowsheet1:K-100	9.59334E+06	160.196	120.061

Flowsheet1:K-100 1	-9.56522E+06	147.778	25.0868	
Flowsheet1:K-101	7.35028E+07	-28.3028	-28.3256	
Flowsheet1:VSSL-100	-1.53626E+07	-44.3740	-59.4672	
Flowsheet1:VSSL-101	-8.01067E+07	-69.4787	-128.042	
Flowsheet1:XCHG-100	-7.45477E+07	147.778	23.4357	

VITA

Ali Abdulhamid Abedi was born in Kuwait City, Kuwait on August 17th 1981. After completing high school in Kuwait, he attended The University of Texas at Austin, where he obtained his B.S. degree in Chemical Engineering in May 2005.

In August 2005, he joined Texas A&M University to begin study toward his M.S. in Chemical Engineering. His research interests are process and production of ethylene, focusing on the new technologies of ethylene production. Mr. Abedi may be reached at the Jack E. Brown Chemical Engineering Building, 1100 Hensel Dr., College Station, Texas, 77840. His email is ali.abedi@chemail.tamu.edu.